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Process Design Manual for Sludge Treatment and Disposal

Black, Crow and Eidsness, Inc, Gainesville, Fla

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FOREWORD

The formation of the United States Environmental Protection Agency marked a new era of environmental awareness in America. This agency's goals are national in scope and encompass broad responsibility in the area of air and water pollution, solid wastes, pesticides, and radiation. A vital part of EPA's national water pollution control effort is the constant development and dissemination of new technology for wastewater treatment.

It is now clear that only the most effective design and operation of wastewater treatment facilities, using the latest available techniques, will be adequate to meet the future water quality objectives and to ensure protection of the nation's waters. It is essential that this new technology be incorporated into the contemporary design of waste treatment facilities to achieve maximum benefit of our pollution control expenditures.

The purpose of this manual is to provide the engineering community and related industry with a new source of information to be used in the planning and design of present and future wastewater treatment facilities. It is recognized that there are a number of design manuals and manuals of standard practice, such as those published by the Water Pollution Control Federation, available in the field that adequately describe and interpret current engineering practices as related to traditional plant design. It is the intent of this manual to supplement this existing body of knowledge by describing new treatment methods, and by discussing the application of new techniques for more effectively removing a broad spectrum of contaminants from wastewater.

Much of the information presented is based on the evaluation and operation of pilot, demonstration, and full-scale plants. The design criteria thus generated represent typical values. These values should be used as a guide and should be tempered with sound engineering judgment based on a complete analysis of the specific application.

This manual is one of several available through the EPA Office of Technology Transfer to describe recent technological advances and new information. Future editions will be issued as warranted by advancing state-of-the-art to include new data as it becomes available, and to revise design criteria as additional full-scale operational information is generated.

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CHAPTER 1

INTRODUCTION

The purpose of this manual is to present a contemporary review of sludge processing technology as well as procedures to be considered, modified, and applied, as appropriate, to meet unique conditions of specific design situations. Current regulations require the design, construction, and cost-effective operation of municipal wastewater treatment plants capable of removing 85 percent or more influent BOD. They further require concurrent isolation and regulated disposal of the resultant sludges. As a result, the task faced by the environmental engineer has become much more complex.

Capital, operating, and maintenance costs of facilities required to provide the higher levels of treatment are significantly greater than those encountered with more elementary levels of treatment. Thus, an increased economic incentive for optimal design exists. In addition to increased quantities of secondary biological sludges, the engineer is now confronted with advanced waste treatment sludges as well as new and more complex industrial wastes. A lack of regulations on sludge disposal procedures has caused the engineering profession to concentrate their design efforts on the liquid treatment portion of the plant. Agencies have not been accustomed to funding concurrent pilot work on sludge processing and in the design of primary plants this was not a serious problem. However, in secondary plants, the various liquid treatment and sludge treatment unit processes are so highly connected and interrelated that both sections must be studied and considered as a unified system. Unit process costs and effectiveness should be quoted in the context of a given system. The fact that the influent waste streams at various cities and localities deviate significantly makes any attempt at standardized design impractical.

The design engineer should have the capability for innovative design. The following items are essential to an optimum innovative design:

1. Development and maintenance of a thorough knowledge of the various sludge treatment unit processes.

2. Continual study of plant operational results to provide feedback for cost saving modifications and future design.

3. Adequate pilot plant study of alternate prescreened treatment plant systems as required by the particular circumstances.

4. Use of the systems analysis method for comparing alternate complete systems (both liquid treatment and sludge processing) for treatment plants.

Case studies and generalized experiences will be discussed in detail to emphasize the numerous sludge treatment and disposal alternatives. Data will be presented to assist the design engineer in selecting the optimum unit processes for inclusion in a particular conceptual design.

The results of inclusion of excess activated sludge and other sludges into processing systems originally designed for primary sludge alone, together with the remedial procedures employed, are also discussed in detail.

CHAPTER 2

METHODOLOGY AND NOMENCLATURE

2.1 Introduction

Suspended solids are usually present in the influents to a municipal wastewater treatment plant at levels of 100 to 300 mg/l. Additional suspended solids are generated during the various wastewater treatment processes. Some result from biological processes and others are generated by chemical precipitation.

Sludge is a broad term used to describe the various aqueous suspensions of solids encountered during treatment. The nature and concentration of the solids control the processing characteristics of the sludge. Grit, screenings, and scum are not normally considered as sludge and, therefore, are not discussed in this manual.

2.2 Methodology

It is important to consider the changing climate or working environment faced by the engineer involved in municipal wastewater plant design. In the past, primary emphasis has been focused on liquid phase wastewater treatment, both in plant design and in research and development work.

Experience indicates that a lack of intense attention to the problems of sludge processing has been prevalent in the engineering profession. As will be noted in chapters 10 and 11, some of the activated sludge plants that came on stream in the United States over the past 5 to 15 years have been plagued with the failure of sludge processing systems to perform as designed, either on a functional or a cost basis, or both.

2.2.1 Working Environment of the Design Engineer

The major cause of the problem just described has been the general climate and specific working conditions regarding design of sludge processing systems facing the engineer. Funds have not been provided in the past for adequate laboratory, pilot plant, and, most important, plant scale process engineering work. This has limited the role of the design engineer in developing innovative techniques. In some cases, responsibility for the design of liquid treatment facilities at a given plant was assigned to one engineering firm while another designed the sludge disposal facilities. This arrangement is not conducive to effective design of an interrelated system.

Problems have plagued the effective design of sludge processing systems despite prior comprehensive and meaningful sludge handling work. Such deficiencies occurred because design was accomplished in a climate that tolerated such actions as periodic wasting of excess solids in plant effluent and conversion of suspended solids into dissolved BOD. With the advent of the Water Quality Act Amendments of 1972, effluent quality has become a prime objective in the design of a wastewater treatment facility. It is now essential to evaluate the effect of all unit processes, such as sludge treatment, on effluent quality.

Current perspectives on the objectives of wastewater treatment plant design include the following facts:

- Both solid and liquid fractions must be satisfactorily processed.
- Capital, operating, and maintenance costs should be optimum for the particular situation.
- Effluent standards are going to be enforced.

2.2.2 Essential Considerations for a Successful Plant

First, optimum conceptual and detailed designs must be prepared. But, since some new effluent standards require new processes, existing textbooks may no longer be adequate. Up-to-date know-how in process engineering and evaluations of plant operating results must be utilized. Second, the plant must be constructed as designed. Third, the plant must be properly operated and maintained after construction. Fourth, continuing plant service and development work should be carried out jointly by the engineer and owner. Plant scale work is the most vital source of information for future process design and modification for improved performance.

2.2.3 The Total System Approach to Design

The main point of perspective in the development of an optimum conceptual design for a wastewater treatment plant is that each unit process must be evaluated as a part of the total system. The most frequently encountered problem in treatment plant design is the tendency to optimize a given subsystem such as sludge dewatering without considering the side effects of this optimization on the overall plant operation and treatment costs. A good example of this is dewatering by vacuum filtration. Many technical articles present an operational analysis including only such factors as dewatering operating costs, production rate, and cake moisture content. To really evaluate a given dewatering process analysis should include, among other things:

- A complete material balance over the dewatering systems.
- The effect of all recycle streams on the operation and cost of other subsystems.

- The ratio of the quantity of volatile solids to the amount of moisture in the filter cake for gauging calorific value.

To illustrate this point, note in Table 2-1 that the percent dry solids level at which autogenous incineration occurs is a function of the calorific value of dry solids in filter cake, which in turn varies with the chemical composition of the solids. The requisite dry solids level for self-sustaining combustion varies from 18.5 percent to 41.8 percent depending on these factors. This is in contrast to the usual claim found in technical articles stating that a 30 to 40 percent dry solids cake is sufficient for autogenous combustion.

TABLE 2-1

AUTHOTHERMIC COMBUSTION [1]

Sludge Parameter	Case 1	Case 2
Gross Calorific Value	17,400	29,100
Percent Combustible Matter in Solids	60	75
Percent Solids	41.8	18.5

Figure 2-1 depicts the diverse array of unit processes and possible sludge treatment schemes for use in modern wastewater plants. This figure further illustrates the marked complexity of these systems.

The total system approach to treatment process selection is based on criteria summarized in Table 2-2 followed by detailed study of the several most likely total plant process systems using the Quantitative Flow Diagram (QFD) method as outlined in Reference [3].

TABLE 2-2

PLANT PROCESS SELECTION CRITERIA

Factors	Considerations
Wastewater Influent	Flow, Characteristics
Liquid Effluent	Standards, Disposal/Utilization
Sludge Treatment	Standards, Disposal/Utilization
Constraints	Existing Facilities Local Environmental–Site Conditions Economics–Capital and O&M Costs Operation (Reliability) Management

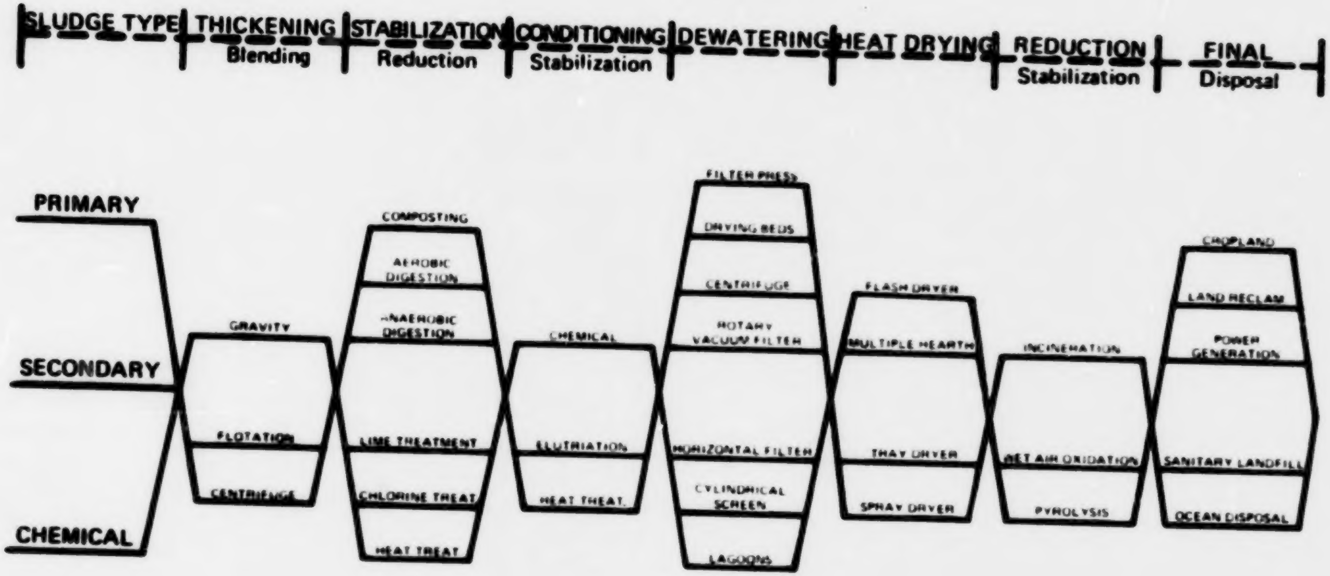

SLUDGE TYPE
THICKENING
Blending
STABILIZATION
Reduction
CONDITIONING
Stabilization
DEWATERING
HEAT DRYING
REDUCTION
Stabilization
FINAL
Disposal
PRIMARY
SECONDARY
CHEMICAL
GRAVITY
FLOTATION
CENTRIFUGE
COMPOSTING
AEROBIC DIGESTION
ANAEROBIC DIGESTION
LIME TREATMENT
CHLORINE TREAT
HEAT TREAT
CHEMICAL
ELUTRIATION
HEAT TREAT.
FILTER PRESS
DRYING BEDS
CENTRIFUGE
ROTARY VACUUM FILTER
HORIZONTAL FILTER
CYLINDRICAL SCREEN
LAGOONS
FLASH DRYER
MULTIPLE HEARTH
TRAY DRYER
SPRAY DRYER
INCINERATION
WET AIR OXIDATION
PYROLYSIS
CROPLAND
LAND RECLAM
POWER GENERATION
SANITARY LANDFILL
OCEAN DISPOSAL

FIGURE 2-1. Unit processes–sludge processing and disposal [2].

The principles involved in the process selection method are further illustrated by Dague et al. [4].

Basic Considerations

In evaluating treatment process alternatives, certain factors should always be considered. Major factors are the wastewater to be treated, effluent requirements to be met, and constraints within which the system must function. Each of the factors listed in Table 2-2 must be considered in relation to all other factors. It is not proper, for example, to consider only the wastewater flow and strength in evaluating a given process. Effluent requirements and other constraints may make the process less attractive.

Once it has been determined that a given system is capable of meeting treatment requirements, within existing geographic and facilities constraints, evaluation is then limited to the three remaining factors in Table 2-2: (1) economic, (2) operational, and (3) management considerations. As with the other elements, these latter factors must be considered in concert. The sludge treatment system with the least dollar cost may not be the best alternative for long-range treatment objectives and this must also be considered. There are elements upon which a dollar value cannot be placed with any degree of reliability. This is especially true in view of rapidly changing environmental requirements. A wastewater system that meets all requirements today may not in the future, should standards change.

Two elements that do not readily appear as dollar costs, although they may contribute to somewhat inflated values, are operational and management problems. The dollar costs to operate and manage one system may be no greater than for another; however, some systems are inherently more reliable than others.

Process Alternatives

It is common practice to classify wastewater treatment processes according to stages of treatment. Methods of handling wastewater are classified as preliminary, primary, secondary, and tertiary treatment. Sludge handling processes can be classified as shown in Figure 2-1. Within each of these categories there are numerous process alternatives.

The challenge in process selection is the evaluation of the numerous system alternatives. A first step in the evaluation is to eliminate inappropriate methods. This first cut is based on considerations listed in Table 2-2. If standards call for a secondary level of wastewater treatment, plus disinfection, then the engineer may elect to eliminate all tertiary treatment methods other than disinfection. Similarly, in sludge processing, when the plant is quite large, the engineer would likely eliminate sand drying beds from detailed economic analysis. If the plant is small, he would probably eliminate incineration as a method of sludge treatment. In short, there are certain alternatives that may be expelled on the basis of engineering judgment. In making the preliminary cut of treatment processes, care must be taken to avoid arbitrary decisions. If the engineer excludes certain treatment methods from further consideration, he should do so only on the basis of sound reasoning, experience, and judgment.

Figure 2-2 illustrates the number of system alternatives that can become involved in analyzing a treatment system. The illustration involves four stages of liquid treatment and two stages of sludge treatment. The result is 432 system alternatives. It is highly impractical for the engineer to prepare a preliminary design and economic analysis for each of these alternatives without the use of a computer. The problem then is one of computer program development. The computer should only be used to make the calculations, not the analysis. It should be used to select the best systems, all within the limits of sensitivity, for the overall analysis. The engineer can then evaluate the more indeterminate aspects of each system and, finally, select the best alternative.

2.2.4 The Design Team Concept

Increased complexity of wastewater treatment plants has led some of those firms specializing in this area to adopt the multidisciplinary approach to plant design. The rationale for this approach is explained in detail by Voysey [5]. It is important from time to time to take a broad look at the science of wastewater treatment, from aspects other than those that are purely technical. It is a science involving many disciplines and the science will develop best if each of those concerned, including operators, consider themselves as members of design teams. This will undoubtedly result in more efficient wastewater treatment. Further design procedures, based on the personal contact between the designer and the operator, will afford improved designs and more efficient operation.

2.3 Sludge Processing and Disposal Nomenclature

To promote understanding and overcome problems of semantics, it is appropriate to review the meaning and significance of certain terms used to describe various types of sludges, processes, and equipment used for their treatment and disposal. The most pertinent publication on wastewater treatment nomenclature is the *Glossary – Water and Wastewater Control Engineering*, prepared by the Joint Editorial Board representing the APHA, ASCE, AWWA, and WPCF. This manual will adhere to that work as closely as possible.

2.3.1 General Considerations

The Quantitative Flow Diagram (QFD) is an important step in the design of a sludge system. This term refers to an engineering process flow sheet which describes the total plant processing system in a quantitative fashion. Such diagrams are essential to analysis of various alternate total plant processing systems. Their use in the subsequent preparation of capacity and cost summations for different combinations of unit processes permits comparison of the cost-effectiveness of such systems.

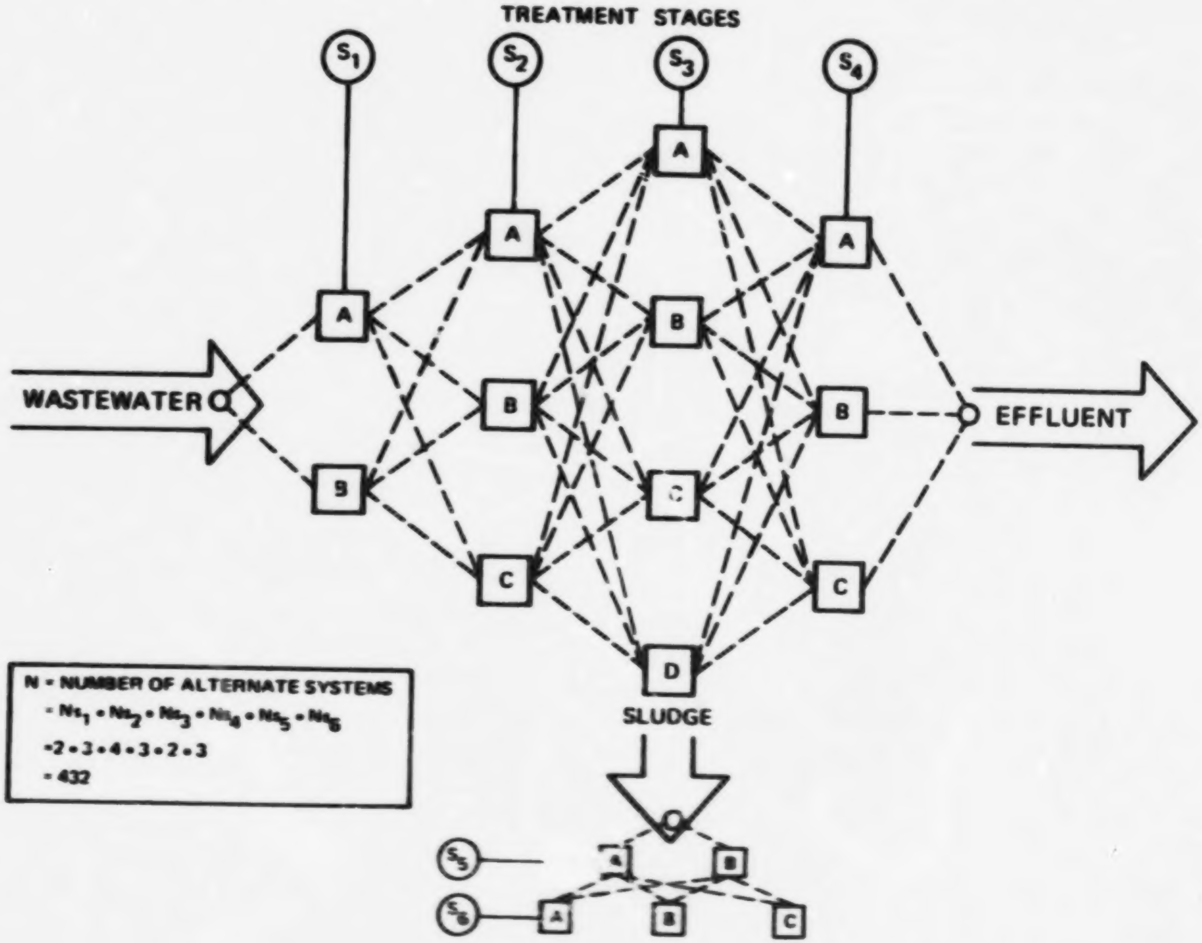

TREATMENT STAGES
S_1
S_2
S_3
S_4
WASTEWATER
EFFLUENT
SLUDGE
A
B
C
D
S_5
S_6
N = NUMBER OF ALTERNATE SYSTEMS
$= N_{S_1} \cdot N_{S_2} \cdot N_{S_3} \cdot N_{S_4} \cdot N_{S_5} \cdot N_{S_6}$
$= 2 \cdot 3 \cdot 4 \cdot 3 \cdot 2 \cdot 3$
$= 432$

FIGURE 2-2. Evaluation of system alternatives.

A material balance is also an essential part of any design. For a detailed explanation the reader is referred to the text by Hougen et al. [6]. A material balance for a process is an exact accounting of all materials which enter, leave, accumulate, or are depleted in the course of a given time of operation. It is, therefore, an expression of the Law of Conservation of Mass. The preparation of an adequate QFD or the execution of process control or development is impossible without consideration and use of material balances. The QFD is essentially a series of interrelated material balances.

Other terms which are relevant to sludge processing are defined as follows:

- *Primary sludge* is the sludge obtained from a primary settling tank. This definition was adequate when all treatment was strictly primary treatment. With the advent of secondary treatment and the recirculation of excess activated sludge and possibly other sludge laden streams to the primary sedimentation basins, the term primary sludge is subject to significant misinterpretation and must be used with care. As originally defined, and in its true perspective, primary sludge is that portion of the raw wastewater solids contained in the raw plant influent which is directly captured and removed in the primary sedimentation process.

- *Biomass* is a synonym for biological solids.

- *Primary sedimentation* is usually the first major process in wastewater treatment works. It is not considered a sludge process. However, in some cases the primary basins are used to capture and thicken sludge.

- *Final sedimentation* is used to some degree in thickening of sludge.

- *Autogenous incineration* refers to the combustion characteristics of a sludge having a composition (physical and chemical) such that no auxiliary fuel is required in incineration (except start-up and shutdown).

2.3.2 Sludge Treatment and Disposal–Unit Processes

The following categorization of processes used in treatment and disposal of sludges is set forth:

- Thickening (Blending)
- Stabilization (Reduction)
- Conditioning (Stabilization)
- Dewatering

- Heat drying
- Reduction (Stabilization)
- Final disposal

In classifying and describing sludge processing methods, the potential of a process to accomplish more than one task must be taken into account. Accordingly, the nomenclature attempts to recognize that four of the major categories (Thickening, Conditioning, Dewatering, and Reduction) have primary as well as secondary objectives.

Sludge Thickening (Blending)

The term thickening, herein, will be used to describe an increase in solids concentration, whether it occurs as the objective of a separate process, or as a secondary effect of a process provided essentially for a different purpose. Thickening Methods (Blending) are as follows:

- Gravity
- Flotation
- Centrifugation

Recognition of the need to uniformly blend or combine the two principal types of wastewater sludges (primary and excess activated), and to keep them combined in plants where joint processing is practiced, is not as widespread as it should be. Normally, sludge blending can best be accomplished in a separate sludge thickening process.

Sludge Stabilization (Reduction)

Sludge stabilization processes are aimed at converting raw (untreated) sludges into a less offensive form with regard to odor, putrescibility rate, and pathogenic organism content. Major types of processes are:

- Anaerobic digestion
- Aerobic digestion
- Lime treatment
- Chlorine oxidation

- **Heat treatment**
- **Composting**

Some discussion of each term follows:

Anaerobic and aerobic digestion involve the biological stabilization of sludge through partial conversion of putrescible matter into liquid, dissolved solids, and gaseous by-products, with some destruction of pathogens. These processes also reduce the amount of dry sludge solids. Consequently, these processes result in stabilization and in solids reduction or conversion.

Lime treatment and chlorine oxidation both control odor and reduce pathogens without significantly reducing sludge solids.

Heat treatment kills pathogenic organisms. In addition, putrescible organic matter is substantially dissolved and appears in the cooking liquor from subsequent decantation or dewatering.

Composting is an aerobic process involving the biological stabilization of sludge. It provides organic solids, pathogen, and odor reduction.

Sludge Conditioning

Sludge conditioning is pretreatment of a sludge to facilitate removal of water in a thickening or dewatering process. Methods are as follows:

- Chemical (Inorganic and Organic)
- Elutriation
- Heat treatment

Chemical methods involve the use of inorganic or organic flocculants to promote formation of a porous, free-draining cake structure. In this way, the flocculants improve sludge dewaterability, alter sludge blanket properties, and improve solids capture. In dewatering, flocculants increase the degree of solids capture both by destabilization and agglomeration of fine particles and facilitate cake formation. The resultant cake becomes the true filter media. In thickening processes, the flocculants promote more rapid phase separation, higher solids contents, and a greater degree of capture.

Elutriation is the process of washing the alkalinity out of anaerobically digested sludge to decrease the demand for acidic chemical conditioners and to improve settling and dewatering characteristics. When used with primary sludge, the process is cost-effective and does not create undesirable effects. When elutriation is used in a plant which combines primary and excess activated sludge prior to digestion, the mixed sludge fractionates during the elutriation process, producing a highly polluted elutriate. The process has been criticized because this elutriate was bypassed into the plant effluent at some plants. However, use of flocculants in elutriation can eliminate the problem of the polluted elutriate.

Heat treatment, herein, refers to the pressure cooking of sludges in such a manner that little sludge oxidation occurs. The Porteous, Farrer, Zurn, and some Zimpro systems fall into this category. Thus, heat treatment is distinct from wet air oxidation which generally involves higher temperatures and pressures, with air injection to promote a major degree of sludge oxidation.

Dewatering Methods

Any process which removes sufficient water from sludge so that its physical form is changed from essentially that of a fluid to that of a damp solid, is a dewatering process. Methods used in dewatering are best described by the equipment employed and some major types are listed below and are discussed in detail in Chapter 7:

- Rotary vacuum filters
- Centrifuges
- Drying beds
- Filter presses
- Horizontal belt filters
- Rotating cylindrical devices
- Lagoons

Heat Drying of Sludge

Sludge drying processes involve the application of heat to evaporate sufficient moisture and render the sludge dry to the touch and relatively free flowing. It is normal practice to conserve energy by dewatering the sludge prior to heat drying. Principal types of dryers are:

- Multiple hearth
- Flash dryers

- Tray dryers
- Spray dryers

Sludge Reduction

Sludge reduction, as defined here, pertains to processes which primarily yield a major reduction in the volatile sludge solids. Principal methods of sludge reduction are:

- Incineration
- Wet air oxidation
- Pyrolysis

Final Disposal Methods

Final or ultimate disposal refers to the disposition of sludge in liquid, cake, dried, or ash form, as a residue to the environment. Principal methods are:

- Cropland application
- Land reclamation
- Power generation (with solid waste)
- Sanitary landfill
- Ocean disposal

The first three methods are also utilization procedures. In instances where sanitary landfills are used for purposes of topographic modification, this also could be construed as utilization.

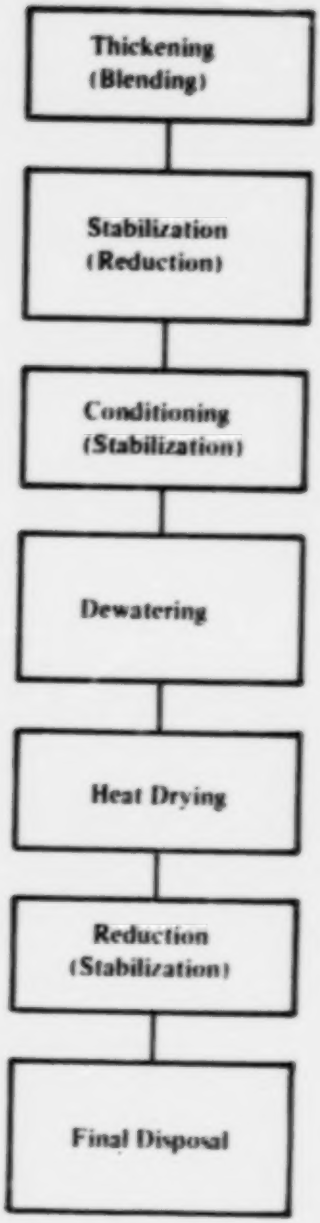
2.3.3 Sequence and Functions of the Unit Processes

Figure 2-3 summarizes the purposes and sequence of unit processes of wastewater sludge treatment.

2.4 References

1. Gale, R. S., "Recent Research on Sludge Dewatering." *Filtr. Separ.* (Sep.-Oct. 1971), pp. 531-538.

UNIT PROCESSES


Thickening (Blending)
Stabilization (Reduction)
Conditioning (Stabilization)
Dewatering
Heat Drying
Reduction (Stabilization)
Final Disposal

FUNCTIONS

- Water Removal
- Volume Reduction
- Post Process Efficiencies
- Blending

- Pathogen Destruction
- Volume and Weight Reduction
- Odor Control
- Putrescibility Control
- Gas Production

- Improve Dewatering or Thickening Rate
- Improve Solids Capture
- Improve Compactability
- Stabilization

- Water Removal
- Volume and Weight Reduction
- Change to Damp Cake
- Reduces Fuel Requirements for Incineration/ Drying

- Water Removal
- Sterilization
- Utilization

- Destruction of Solids
- Water Removal
- Conversion
- Sterilization

- Utilization (Cropland)
- Utilization (Energy)
- Utilization (Land Reclamation)
- Disposal (Landfill)
- Disposal (Ocean)

FIGURE 2-3. Enumeration of sludge treatment processes and their functions.

2. Stanley Consultants, "Sludge Handling and Disposal, Phase I–State of the Art." Report to Metropolitan Sewer Board of the Twin Cities Area, Nov. 15, 1972.

3. Camp, Dresser & McKee, Inc., "Municipal Wastewater Treatment Plant Sludge and Liquid Sidestreams." Preliminary Technical Bulletin, for the EPA, Contract 68-01-0324, 1973.

4. Dague, R. R., Walker, J. T., and Moritz, P. J., "Evaluation of Treatment Process Alternatives: Two Case Studies." Presented at the 17th Great Plains Wastewater Design Conference, Omaha, Nebraska, Mar. 27, 1973.

5. Voysey, J. A., "Overall Design Procedures, with Particular Reference to Contact Between Designer and Operator." *Water Pollut. Contr.* (1973), pp. 231-234.

6. Hougen, O. A., Watson, K. M., and Ragatz, R. A.,*Chemical Process Principles* (2nd ed., pt. 1). John Wiley & Sons: New York (1954).

CHAPTER 3

OCCURRENCE OF SLUDGES AND PHYSICAL AND CHEMICAL PROPERTIES RELATING TO PROCESSABILITY

3.1 Occurrence of Sludges–Conventional Biological Treatment

The various wastewater treatment processes produce different amounts and types of sludges. Table 3-1 presents data on typical volumes of sludges produced in several conventional treatment processes.

TABLE 3-1

TYPICAL SLUDGE VOLUMES PRODUCED

Wastewater Treatment Process	Gallons of Sludge Produced/Million Gallons Wastewater Treated			
	Keefer [1]	Fair & Imhoff [2]	Babbitt [3]	McCabe and Eckenfelder [4]
Primary sedimentation	2,950	3,530	2,440	3,000
Trickling filter	745	530	750	700
Activated sludge	19,400	14,600	18,700	19,400

This table illustrates the striking increase in the volume of sludges to be processed when a plant is upgraded to activated sludge treatment.

Table 3-2 illustrates typical masses or weights of sludges produced by various conventional methods of treatment.

TABLE 3-2

TYPICAL SLUDGE MASSES [5]

Wastewater Treatment Process	Percent Suspended Solids Removed by Process	Pounds of Solids Generated by Process per Million Gallons Treated	Specific Gravity of Suspended Solids
Primary sedimentation	60	1,020	1.33
Trickling filter	30	510	1.52
Primary plus activated sludge	92	1,563	1.33

Comparison of Tables 3-1 and 3-2 shows that of the three most common sludge types, excess activated sludge is by far the most voluminous. The large volume is due primarily to its low solids concentration. This point is further illustrated in Table 3-3.

TABLE 3-3

TYPICAL WATER CONTENT OF SLUDGES [6]

Wastewater Treatment Process	Percent Moisture of Sludge Generated	lb Water/lb Sludge Solids
Primary sedimentation	95	19
Trickling filter		
Humus–low rate	93	13.3
Humus–high rate	97	32.3
Activated sludge	99	99

While the data in Tables 3-1, 3-2, and 3-3 are typical, consideration must still be given to the analysis of a particular wastewater and process efficiency in order to accurately estimate sludge quantities.

3.2 Occurrence of Sludges–Combined Biological and Chemical Treatment

Addition of metallic salts at various points within conventional biological treatment systems, for upgrading levels of treatment or phosphorus removal, results in the production of additional quantities of sludges. Increased amounts of sludge result primarily from precipitation of insoluble metallic compounds, largely phosphates. Table 3-4 delineates an estimation procedure for calculating the increase in sludge mass when iron or aluminum are added at various points in the wastewater treatment sequence.

TABLE 3-4

SAMPLE CALCULATION FOR ESTIMATING SLUDGE MASS (lb/M.G.) [7]

	Conventional	Fe to Primary	Fe to Aerator	Al to Aerator	Al to TF Clarifier
Primary					
SS removal	50%	75%	50%	50%	50%
Sludge solids	1,250	1,875	1,250	1,250	1,250
Fe solids	0	605			
Al solids	0				
Total	1,250	2,480	1,250	1,250	1,250
Activated Sludge					
Secondary solids	715	536	804	804	
Fe solids			541		
Al solids				425	
Total (System)	1,965	3,016	2,595	2,479	
Trickling Filter					
Secondary solids	656				745
Al solids					483
Total (System)	1,906				2,478

Basis for Sludge Mass Calculation

Cation/P Dose (mol/mol)	lb Chemical Sludge/lb Cation	
	lb/lb Al	lb/lb Fe
1.5	3.9	2.4
1.75	3.8	2.3

Assumptions

Cation/P Dose = 1.5 mol/mol to aerator
Cation/P Dose = 1.75 mol/mol to primary or before trickling filter clarifier
Influent Wastewater
BOD = 230 mg/l
SS = 300 mg/l
P = 10 mg/l

The quantity of sludge resulting from the use of lime in either primary or tertiary treatment can be estimated from wastewater analyses as shown in Table 3-5.

TABLE 3-5

SAMPLE CALCULATION OF SLUDGE QUANTITY FROM LIME TREATMENT OF WASTEWATER [8]

Lime Added Primary Treatment

Data Required:

On influent and effluent: alkalinity, pH, calcium hardness, phosphorus.

Change in Ionic Content (Influent - Effluent)	mg/l
HCO_3, as $CaCO_3$	223
CO_2, as $CaCO_3$	14
Mg, as $CaCO_3$	66

Sludge Produced	mg/l
hydroxyapatite	27
$CaCO_3$	460
$Mg(OH)_2$	38
Total Calculated Sludge	525

Meas./Calculated 1.25

Material Balance on Ca

$Ca(OH)_2$ dose = 390 mg/l
Input-Output = 2.9 mg/l

Following is a recent review of thirteen actual case studies giving the sludge production data on primary, secondary, and tertiary phosphorus removal systems [7].

Table 3-6 illustrates the additional sludge to be processed when chemicals are added to the primary tanks for phosphorus removal.

TABLE 3-6

ADDITIONAL SLUDGE TO BE HANDLED WITH CHEMICAL TREATMENT SYSTEMS: PRIMARY TREATMENT FOR REMOVAL OF PHOSPHORUS

Sludge Production Parameter		Conventional Primary	Low Lime Addition to Primary Influent	High Lime Addition to Primary Influent	Aluminum Addition to Primary Influent	Iron Addition to Primary Influent
Level of chemical addition (mg/l)		0	350-500	800-1,600	13-22.7	25.8
Percent sludge solids	Mean	5.25	11.1	4.4	1.2	2.25
	Range	5.0-5.5	3.0-19.5	2.1-5.5	0.4-2.0	1.0-4.5
lb/M.G.	Mean	788	5,630	9,567	1,323	2,775
	Range	600-950	2,500-8,000	4,700-15,000	1,200-1,545	1,400-4,500
gal/M.G.	Mean	4,465	8,924	28,254	23,000	21,922
	Range	3,600-5,000	4,663-18,000	16,787-38,000	10,000-36,000	9,000-38,000

Table 3-7 shows similar data for chemical additions to the secondary system for phosphorus removal.

TABLE 3-7

ADDITIONAL SLUDGE TO BE HANDLED WITH CHEMICAL TREATMENT SYSTEMS: PHOSPHORUS REMOVAL BY MINERAL ADDITION TO AERATOR

Sludge Production Parameter		Al^{+++} Addition to Aerator		Fe^{+++} Addition to Aerator	
		Conventional Secondary	With Al^{+++} Addition	Conventional Secondary	With Fe^{+++} Addition
Level of chemical addition (mg/l)		0	9.4-23	0	10-30
Percent sludge solids	Mean	0.91	1.12	1.2	1.3
	Range	0.58-1.4	0.75-2.0	1.0-1.4	1.0-2.2
lb/M.G.	Mean	672	1,180	1,059	1,705
	Range	384-820	744-1,462	918-1,200	1,100-2,035
gal/M.G.	Mean	9,100	13,477	10,650	18,650
	Range	7,250-12,300	7,360-20,000	10,300-11,000	6,000-24,000

Data in Table 3-8 indicate the additional sludge resulting from tertiary phosphorus removal with chemicals.

TABLE 3-8

ADDITIONAL SLUDGE TO BE HANDLED WITH CHEMICAL TREATMENT SYSTEMS: PHOSPHORUS REMOVAL BY MINERAL ADDITION TO SECONDARY EFFLUENT

Sludge Production Parameters		Lime Addition	Aluminum Addition	Iron Addition
Level of chemical addition (mg/l)	--	268-450	16	10-30
Percent sludge solids	Mean	1.1	2.0	0.29
	Range	0.6-1.72	--	--
lb/M.G.	Mean	4,650	2,000	507
	Range	3,100-6,800	--	175-781
gal/M.G.	Mean	53,400	12,000	22,066
	Range	50,000-63,000	--	6,000-36,000

In summation:

- Lime addition in the primary causes the greatest increase in sludge mass production.
- The minimum increase in sludge mass comes from use of alum in the aeration basins.
- Sludge mass and volume depend critically on wastewater characteristics and clarifier performance.
- Selection of a treatment chemical and point of application should take into account the relative sludge processability as well as added sludge mass.

3.3 Physical and Chemical Properties Relating to Processability

This section relates the measurable physical and chemical properties of types of sludges to their processing characteristics. Further information on sludge conditioning can be found in Chapter 6.

3.3.1 Factors Affecting Processing

The sludge characteristics affecting processability are:

- Particle density
- Particle size distribution
- Surface charge
- Degree of hydration
- Compressibility (blanket and cakes)

Particle density is most important in sludge thickening since it affects the subsidence rate. Particle size distribution affects both thickening and dewatering. As the average particle size decreases, the surface/volume ratio increases exponentially. Increased surface area means greater hydration, higher chemical demand, and increased resistance to dewatering. Particle size distribution also affects compressibility.

The two principal factors promoting stable dispersions of solids in liquids are surface charge and hydration. Surface charge must be neutralized or circumvented to promote flocculation. Regardless of surface charge, other surface effects can cause solid particles to adsorb film layers of water (hydration) and this also can adversely affect moisture removal processes. Wastewater sludges all contain dispersed solids with some degree of surface charge and hydration. It is an unfortunate fact that solids contained in essentially all municipal wastewater sludges yield compressible and hydrophilic sludge cakes unless the sludge prior to dewatering has been coagulated or flocculated. Cake compressibility and migration of unstructured fines inhibits water removal in thickening and dewatering.

3.3.2 Particle Size and Configuration

Figure 3-1 shows the relative particle size of common materials. Fines and colloidal particles are present in raw wastewater and practically all escape capture in primary basins. The activated sludge process, in addition to removal of dissolved BOD, functions to capture, remove, and partially metabolize these materials. Unfortunately, some of the various wastewater processes, both liquid and solid, tend to create additional fines. The other major processes for removing fines are coagulation and flocculation. In those rare instances where some form of treatment is not required to promote thickening and dewatering, relatively unhydrated materials abound in the sludge solids. Table 3-9 depicts the relative difficulty of removing water from an unflocculated primary digested sludge containing various particle size fractions.

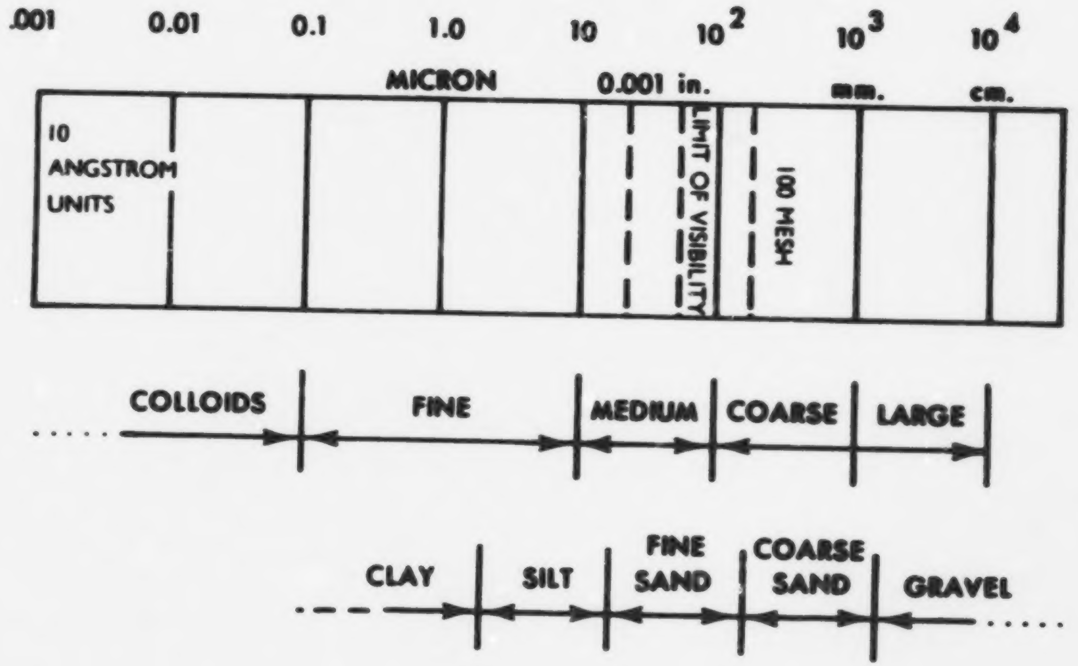

.001
0.01
0.1
1.0
10
10^2
10^3
10^4
MICRON
0.001 in.
mm.
cm.
10 ANGSTROM UNITS
LIMIT OF VISIBILITY
100 MESH
COLLOIDS
FINE
MEDIUM
COARSE
LARGE
CLAY
SILT
FINE SAND
COARSE SAND
GRAVEL

FIGURE 3-1. Particle size common materials.

TABLE 3-9

SLUDGE DEWATERING AS A FUNCTION OF PARTICLE SIZE [9]

Mean Diameter (microns)	Specific Resistance (sec^2/g)	Coefficient Compressibility	Percent of Total Particles
Original, unfractionated sample	10.4×10^9	0.66	–
Above 100	2.3×10^9	0.73	10.2
5-100	4.6×10^9	0.70	75.5
1-5	13.8×10^9	0.42	8.5
Below 1	–	–	5.9

As can be seen, the specific resistance to filtration of the unfractionated sludge is dominated by the specific resistance of material under 5 microns in size, even though this material constitutes only about 14 percent by weight of the total solids. Specific resistance is in effect, a measure of the relative dewaterability of a sludge. It has been defined as the pressure required to produce a unit rate of flow through a cake having a unit weight of dry solids per unit area when the viscosity of the liquid is unity. Specific values are determined from laboratory filtration experiments. The seeming paradox on the coefficient of compressibility is tempered by the facts that:

- The significant compressibility is that of the unfractionated cake due to interactions of the fractions.
- All of the fractions are sufficiently compressible to prevent unflocculated dewatering.
- Greater compressiblity of coarser particles may result from some of these coarser particles being agglomerates of smaller particles. In any event, the much lower specific resistance of the coarser particles completely overshadows the compressibility difference. Apparently, the water retention characteristics of the fines must be the dominant factor.

Table 3-10 depicts typical specific resistance values for different types of sludges, both coagulated and uncoagulated. Since the maximum specific resistance for feasible mechanical dewatering is normally quoted at 1.0×10^8 (sec^2/g), none of these sludges would be readily dewaterable. Table 3-10 shows that specific resistance values can vary significantly. The actual value for a particular sludge will be a function of both its origin and nature. Broad experience indicates that properly conditioned raw primary sludge is almost always the most readily dewaterable, followed by well-conditioned, digested, and activated sludges in increasing order of difficulty.

TABLE 3-10

SPECIFIC RESISTANCE OF VARIOUS TYPE SLUDGES [9]

Type Sludge	Specific Resistance (sec^2/g)
Raw	$10\text{-}30 \times 10^9$
Raw (coagulated)	$3\text{-}10 \times 10^7$
Digested	$3\text{-}30 \times 10^9$
Digested (coagulated)	$2\text{-}20 \times 10^7$
Activated	$4\text{-}12 \times 10^9$

3.3.3 Surface Charge and Degree of Hydration

Both thickening and dewatering require aggregation or packing of individual particles into larger clumps. Destabilization or flocculation occurs only after the dispersed particles' normal net negative charge is essentially neutralized and some degree of surface water desorption and particulate bridging has been brought about.

The desirable changes are brought about in three ways:

- Natural flocculation in and by activated sludge organisms through the mechanism of excretion of natural polymeric materials.
- Use of synthetic organic polymer flocculants.
- Use of inorganic metal salts as primary coagulants.

The amount of polymeric flocculants excreted by microorganisms in the activated sludge process is directly related to the length of aeration. In turn, the settleability of activated sludge solids is a function of the natural polymer present and this is illustrated in Table 3-11.

TABLE 3-11

VARIATION OF SVI WITH AERATION TIME [10]

Process Variation	Sludge Volume Index	Aeration Period (hr)
High rate activated sludge	130-480	3-6
Conventional activated sludge	80-130	6-7.5
Extended aeration	50-80	16-24

Table 3-11 shows that increased aeration times yield a more readily settleable excess activated sludge, as measured by the sludge volume index (SVI). These results should not be used in a dogmatic fashion. Recent evidence shows an optimum aeration period usually exists for a given wastewater treatment system and prolonged aeration and overproduction of naturally generated polymer can be detrimental. Pilot scale work is still the best determinant.

Additional insight into the effect of processing variations is illustrated by the data in Table 3-12.

TABLE 3-12

EFFECT OF STORAGE TIME ON SLUDGE DEWATERABILITY [11]

Storage Time (days)	Temperature (°F)	Dewatering (gpm)	Chemical Dosage (mg/l)
8	40	22	150
None	90-100	66	100

These tests show that storage of a digested mixed sludge can have a drastic effect on dewaterability. These results were not significantly affected by temperature.

Estimated distribution of aqueous fluid in a digested mixed sludge is shown in Table 3-13.

TABLE 3-13

AQUEOUS FLUID DISTRIBUTION IN DIGESTED SLUDGE [12]

	Percent Water	Percent Solids
Between cell water	70	5
Adhesion and capillary water	22	15
Adsorption and intracellular fluid	8	40

Generally, thickening removes most of the first type of water shown in Table 3-13. Dewatering can remove most of the adhering and capillary water yielding a solids content of 20 to 40 percent. Further dehydration requires drying or combustion.

3.3.4 Compressibility and Water Retention

Filtration is the operation of separating a heterogeneous mixture of fluid and solid particles by means of a filter medium which permits the passage of the fluid but retains most of the particles. Personnel engaged in work on wastewater sludge dewatering should be familiar with the facts that cake filtration is the prevalent separation mechanism and wastewater sludge cakes are practically always compressible and hydrophilic unless the sludge has been previously conditioned. Cake formation rate, porosity, structure, and water release characteristics are the dominant factors in cake filtration, not the open space or tightness of the media. Figures 3-2 and 3-3 show the effect of compressibility on the pressure drop through incompressible and compressible filter cakes.

In the idealized incompressible cake (see Figure 3-2) the voidage is constant throughout; even though a pressure gradient exists, the particles do not deform or decrease porosity. There is no dramatic increase in resistance to filtration as a function of time. In the case of the compressible cake (see Figure 3-3) the particle deformation will decrease voidage and thus increase resistance to filtration. Degree and rate of deformation, hence resistance to filtration, will be more pronounced at the bottom of the cake. Compressibility and water retention characteristics of wastewater sludge cakes are materially decreased by the various conditioning methods to facilitate dewatering. This is because conditioning procedures eliminate or agglomerate fine particles to minimize their decreasing cake voidage and drainability.

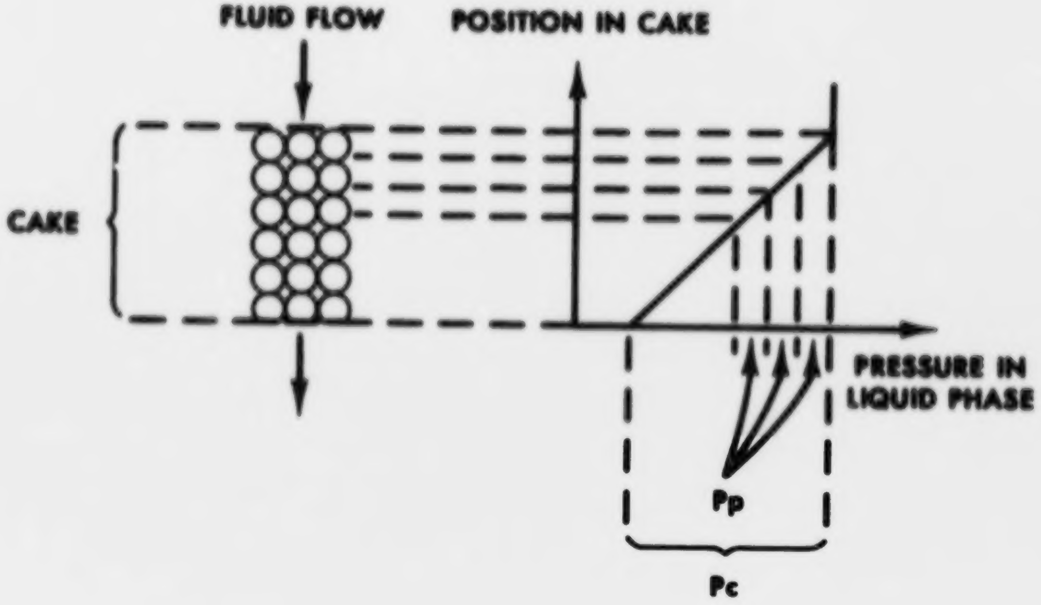

FLUID FLOW
POSITION IN CAKE
CAKE
PRESSURE IN LIQUID PHASE
Pp
Pc

FIGURE 3-2. Fluid flow through an incompressible sludge cake [13].

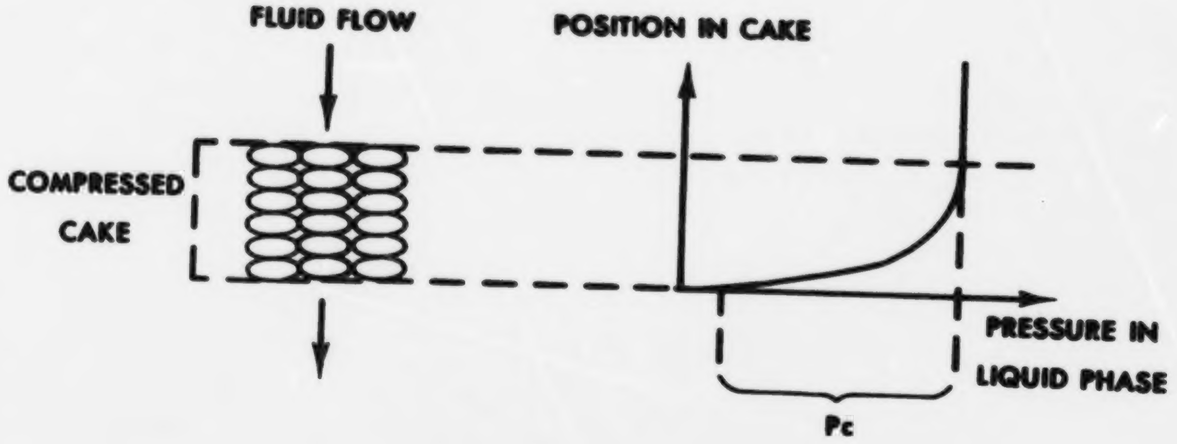

FLUID FLOW
POSITION IN CAKE
COMPRESSED CAKE
PRESSURE IN LIQUID PHASE
Pc

FIGURE 3-3. Fluid flow through a compressible sludge cake [13].

3.4 Plant Experiences with Various Processes and Types of Sludge

This section relates experience on specific types of sludges and processability. In the case of sludge dewatering, additional material may be found in Chapter 7.

3.4.1 Raw Primary Sludge

The nature of primary sludge resulting from the initial clarification step varies to some degree with the makeup of the collection system, ambient temperature, and relative amounts and types of industrial wastes included. However, in nearly all cases, primary sludge is fairly coarse and fibrous in nature, compacts well in thickening processes, dewaters easily, and facilitates autogenous incineration when that method of solids reduction is employed. Figure 3-4 is a close-up photo of primary filter cake.

Thickening of raw primary sludges, preferably carried out immediately in a gravity thickener, is usually an efficient and simple process. Significant savings usually result in subsequent processes such as digestion and dewatering.

Figure 3-5 illustrates the appearance of raw primary sludge cake during dewatering on a vacuum filter belt. Note the excellent release from the filter belt. The heavy, thick, and relatively dry filter cake affords a significant degree of ease in the operation.

As indicated in Table 3-14, dewatering of thickened raw primary sludge by vacuum filtration is an efficient unit operation. It gives high yields at low costs and excellent solids capture prevents recirculation load problems.

TABLE 3-14

TYPICAL PERFORMANCE DATA FOR THE VACUUM FILTRATION OF RAW PRIMARY SLUDGE

Conditioner Used	Percent Sludge Solids	Chemical Cost ($/ton)*	Yield (lb/ft²/hr)	Cake Solid (%)	Solids Capture (%)
Cationic polymer	10	1.67	10	32	90-95

*Based on 1971 cost figures.

FIGURE 3-4. Raw primary filter cake.

Reproduced from
best available copy.

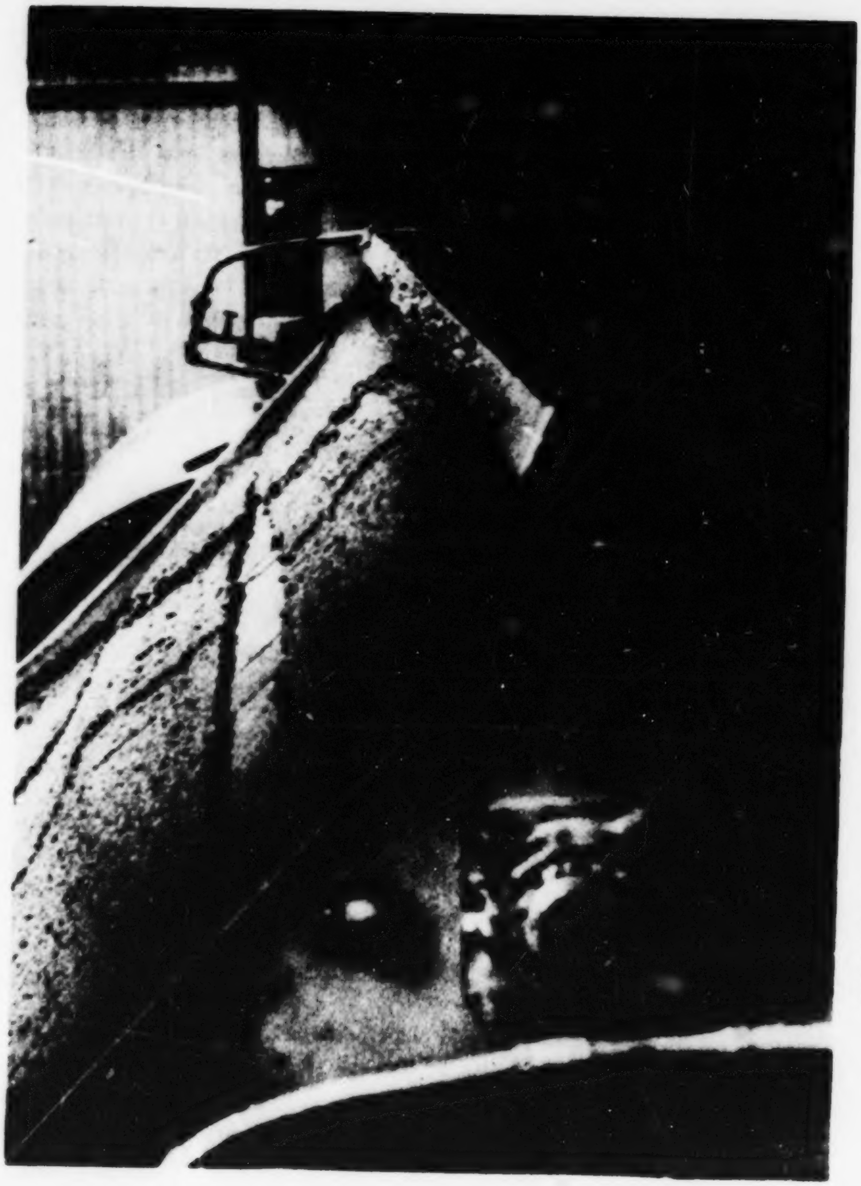


FIGURE 3-5. Removal of raw primary sludge cake from a vacuum filter.

While primary sludges do give somewhat compressible filter cakes, they are easy to condition with polyelectrolytes because of sufficient gross solids (~ 30 percent > 30 mesh). These solids permit rapid formation of filter cake, an adequate structural matrix to provide excellent solids capture, and a rapid dewatering. Rapid cake forming tendency of conditioned raw primary sludge also affords use of relatively coarse filter media with a high percentage of solids capture.

3.4.2 Effect of Anaerobic Digestion on Primary Sludge Dewatering

Handling characteristics of primary sludges are not greatly changed by either standard or high-rate digestion processes. The sludge still settles, compacts, and dewaters easily. However, experience in many plants does indicate that digestion yields a sludge which is slightly more difficult to dewater than raw primary sludge. Typical operating data are presented in Table 3-15. When compared to data in Table 3-14, costs for filtering digested sludge are higher and yields poorer than for raw primary sludge. Operating characteristics during dewatering are about the same with relatively trouble-free operation being normal.

TABLE 3-15

TYPICAL PERFORMANCE DATA FOR
VACUUM FILTRATION OF DIGESTED PRIMARY SLUDGE

Percent Sludge Solids	Conditioner Cost ($/ton)*	Yield (lb/ft²/hr)	Cake Solids (%)	Solids Capture (%)
12.7	2.64	7.4	28	90+

*Based on 1971 cost figures.

3.4.3 Activated Sludge from Conventional Air Systems

Conventional air systems yield activated sludges that are inherently more variable in nature than primary sludge. The principal source of variation is the configuration and mode of operation of the particular system. However, it is possible to generalize, within limits, on the nature and handling characteristics of biomass from essentially domestic sanitary waste.

Activated sludge is finer in particle size than primary sludge. It is normally comprised of 60 percent to 90 percent or more cellular organic material and contains a very large amount of water. Elemental particles of activated sludge are usually aggregated to some extent through bioflocculation. The settling process for activated sludge is hindered because of interparticle interferences and fluid forces resulting in occurrence of zone settling. Further, zone settling rate and compaction tendency are sensitive to the degree of flocculation of the activated sludge. Accordingly, the degree of flocculation of activated sludge achieved, in either an aeration basin via bioflocculation or by the use of flocculants, will affect the settling characteristics as well as the degree of solids capture.

Tenney, Echelberger, Coffey, and McAloon [14] illustrated the effect of biokinetic parameters on settling and dewatering properties of resultant activated sludge. Figure 3-6 illustrates the marked effect of the microogranism growth phase on the percent dispersion or conversely, the degree of bioflocculation occurring.

Concurrent data on the amount of natural biopolymer produced by the microorganisms and consequent effects on dewaterability of the activated sludge are shown in Figure 3-7.

These figures show that processing characteristics of excess activated sludge and the amount which has to be processed are functions of the particular treatment plant's operation. While Tenney et al. [14] indicate that prolonged aeration and the associated accumulation of natural biopolymer result in more efficient drainability, the work by Randall et al. [15] shows that an optimum aeration time and biopolymer accumulation occur in about 120 hours of aeration. Further aeration worsens drainability. Though foregoing examples show the beneficial effect of a long aeration period, plants do not normally operate activated sludge systems in such a manner that significant aerobic digestion and improved processing characteristics result. Additional readings on the subject of biopolymer effects on activated sludges are listed at the end of this chapter.

Biomass tends to float rather than sink after being subjected to shear in various handling processes such as pumping. The degree of bioflocculation acheived can also help considerably in overcoming effects of shear during handling. In some cases, well-flocculated aggregates of activated sludge have been identified even after anaerobic digestion and elutriation.

Experiences have shown that gravity thickening of activated sludge is not a practical operation, particularly if the sludge has been exposed to shear and is not well bioflocculated. Biomass, however, is very amenable to flotation thickening and production of a 4 to 6 percent solids sludge is routine. It should also be noted that the rheological properties of thickened activated sludge are much different from those of primary sludge.

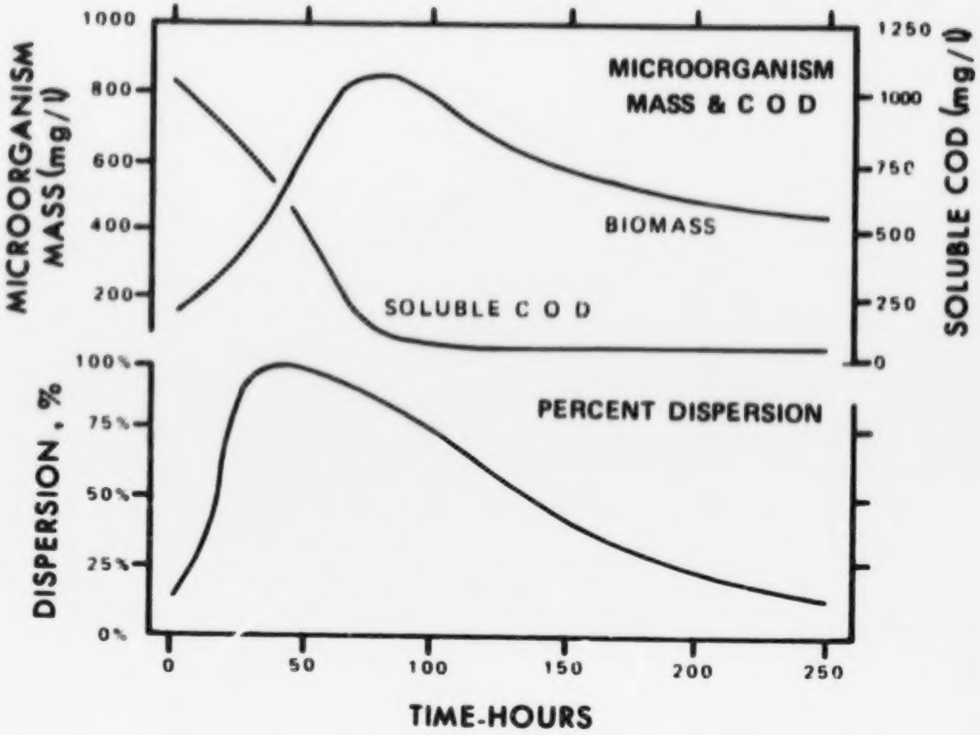


FIGURE 3-6. Effect of time on microorganism mass, COD, and percent particle dispersion [14].

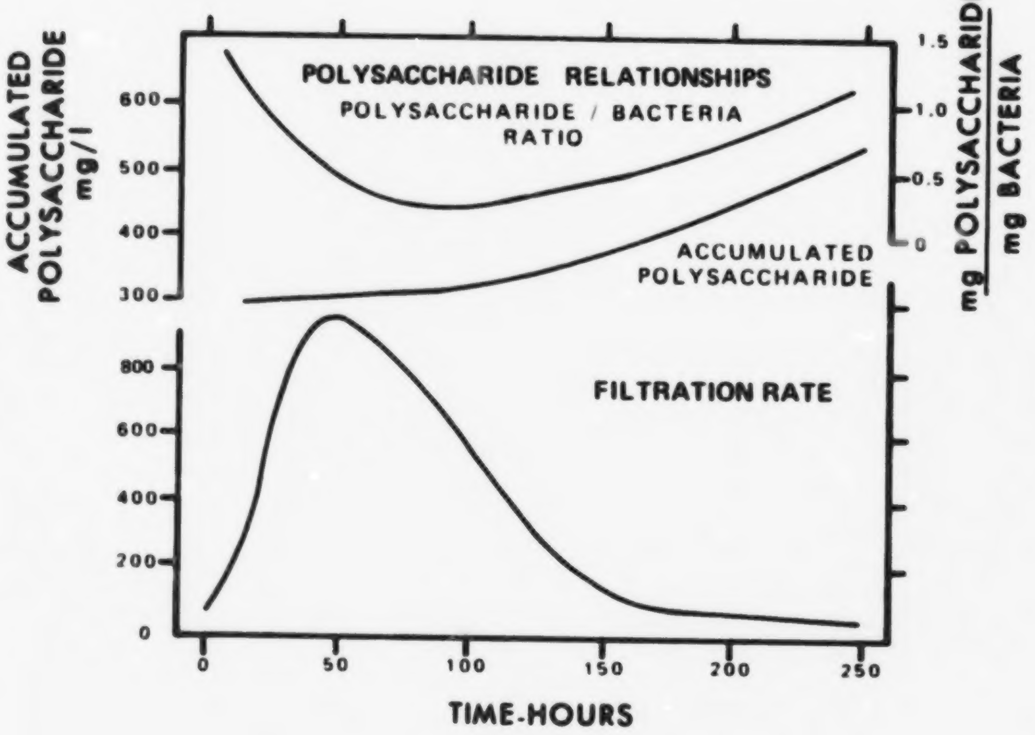


FIGURE 3-7. Effect of time on biopolymer accumulation and dewaterability [14].

3.4.4 Rationale of Design for Some Existing Activated Sludge Plants

Previous observations on nature and handling characteristics of primary sludges and excess activated sludges are important to the design and successful operation of plants combining the two distinctly different sludges early in the solids handling process. In many cases, plants have been designed and constructed to recirculate excess activated sludge to the head of the plant and mix it with primary solids in the primary clarifier. The mixed sludge is then processed in a manner analogous to methods previously described for straight primary sludges. Figure 3-8 illustrates such a design.

Since settling and compaction properties of the excess activated sludge as previously described prevail and dominate throughout the system, processing problems occur. Solids capture in primary basins almost always degrades due to recirculation of activated sludge. This results in a greater load on the secondary system and production of more excess sludge. The combined sludges after anaerobic digestion do not settle well, and this results in a high solids supernatant and a serious solids recirculation problem. When subjected to elutriation, the combined sludge fractionates with resultant poor solids capture and an additional solids recirculation problem. Further, the combined sludges do not compact well without elutriation and this means the dewatering operation must handle a low solids concentration sludge at high costs and low yields.

In other cases plants have been designed so that the excess activated sludge is mixed with primary sludge just prior to gravity thickening, digestion, elutriation, and dewatering. Such a plant is illustrated in Figure 3-9. Though the inclusion of an early thickening step aids primary digester operation, subsequent solids processing steps are still plagued with the same problems previously described. Difficulties experienced in plants which operate in this manner are described more fully in the case studies presented in Chapter 10. In these case studies, remedial measures were taken to make the sludge systems function. Different conceptual designs could give lower costs.

3.4.5 Improvements in Mixed Sludge Processing

The design of air activated sludge systems to provide some degree of endogenous respiration and improved sludge processability was discussed previously. It is generally beneficial to keep air activated sludge out of the primary sludge processing stream until just prior to the dewatering step. Assuming normal primary sludge quality and yield, and excess air activated sludge in normal ratio and quality, the system depicted in Figure 3-10 would appear to be advantageous when raw dewatered sludge is acceptable for final disposal.

This design has the advantage of a separate thickener for each of the sludges which after blending provides the filter with a high solids concentration sludge. Efficient dewatering is promoted with minimization of the deleterious effect of the heavy spiralling

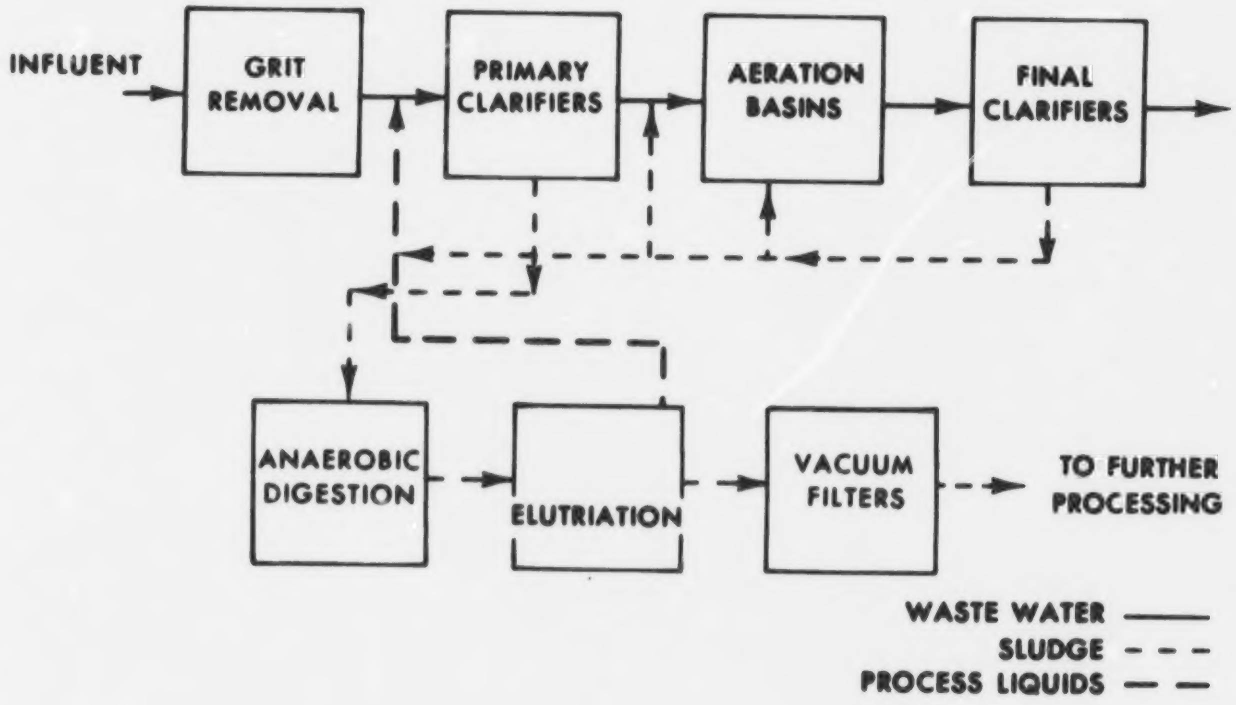

INFLUENT
GRIT REMOVAL
PRIMARY CLARIFIERS
AERATION BASINS
FINAL CLARIFIERS
ANAEROBIC DIGESTION
ELUTRIATION
VACUUM FILTERS
TO FURTHER PROCESSING
WASTE WATER ——
SLUDGE - - -
PROCESS LIQUIDS — —

FIGURE 3-8. Activated sludge plant where EAS is recirculated to primary clarifiers.

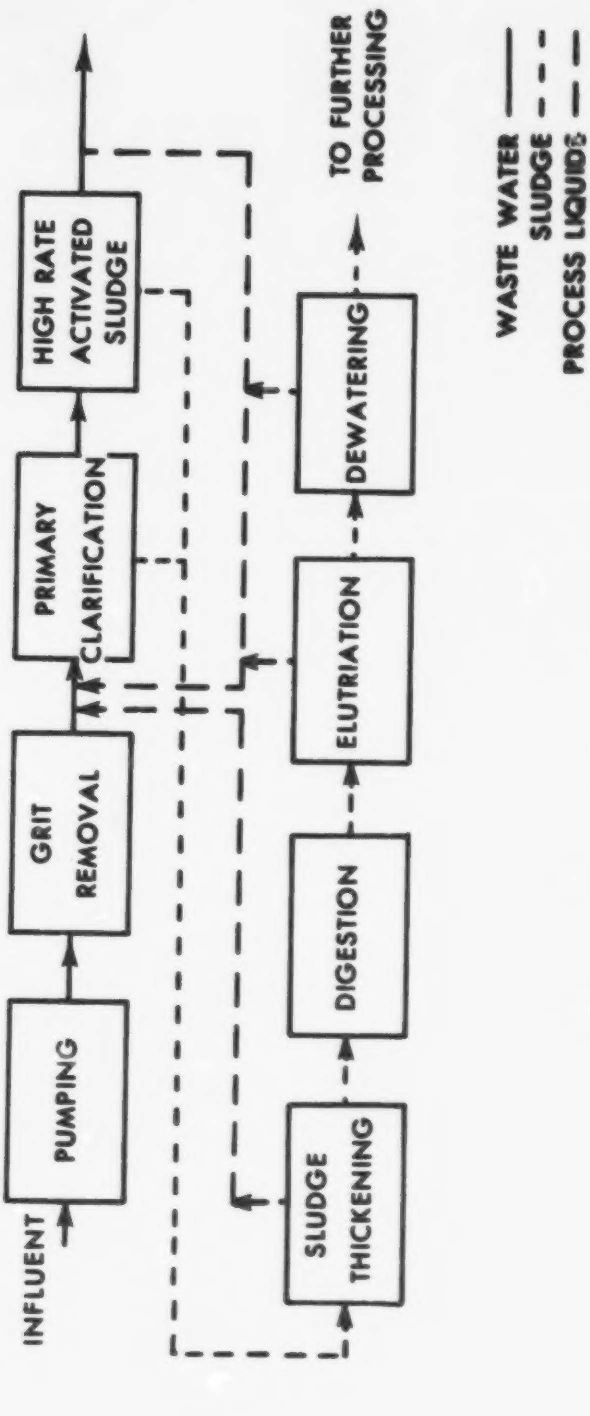

INFLUENT
PUMPING
GRIT REMOVAL
PRIMARY CLARIFICATION
HIGH RATE ACTIVATED SLUDGE
SLUDGE THICKENING
DIGESTION
ELUTRIATION
DEWATERING
TO FURTHER PROCESSING
WASTE WATER
SLUDGE
PROCESS LIQUIDS

FIGURE 3-9. Activated sludge plant where EAS is mixed with primary sludge prior to thickening and digestion.

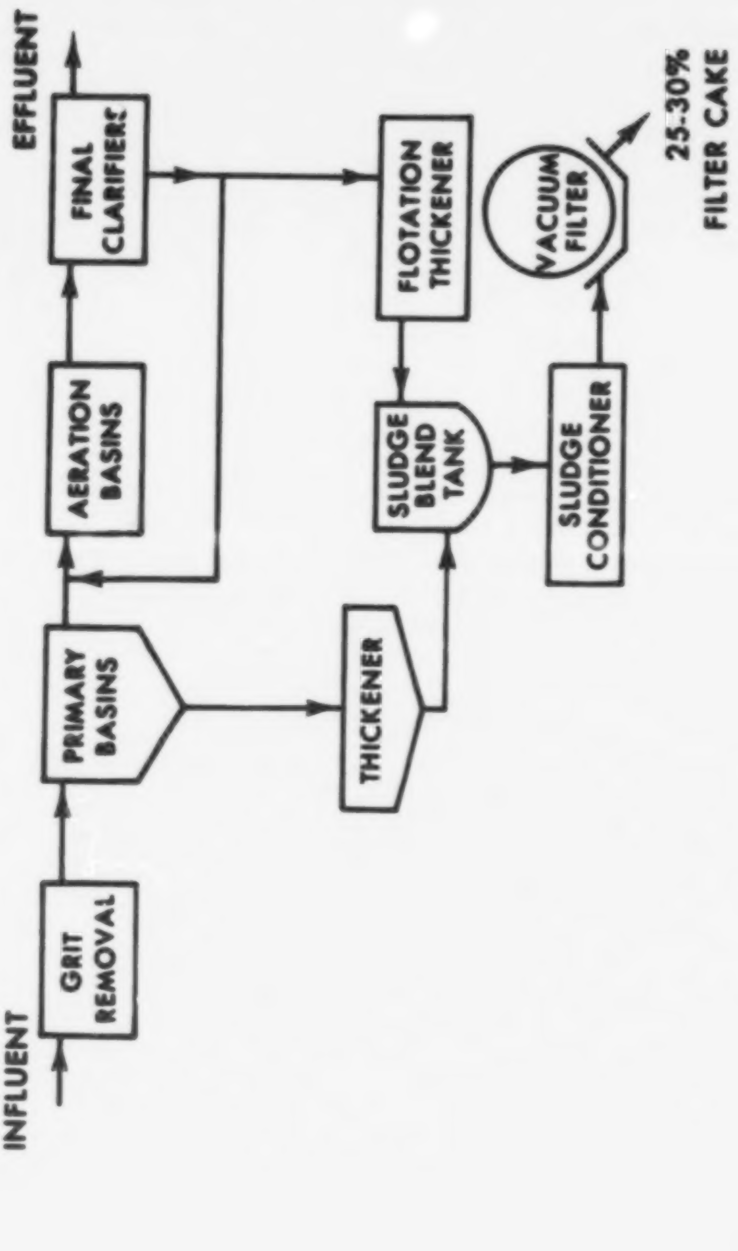

INFLUENT
GRIT REMOVAL
PRIMARY BASINS
AERATION BASINS
FINAL CLARIFIERS
EFFLUENT
THICKENER
FLOTATION THICKENER
SLUDGE BLEND TANK
SLUDGE CONDITIONER
VACUUM FILTER
25-30% FILTER CAKE

FIGURE 3-10. Improved sludge processing scheme for an activated sludge plant.

recirculation loads associated with the two previously mentioned design schemes. The provision of separate flotation thickening of excess air system activated sludge is necessary to adequately thicken this material on a reliable basis.

In the event that higher than normal ratios of air system activated sludges must be processed, inclusion of aerobic digestion should be considered for reducing the organic solids. Post aerobic digestion thickening is usually required and accomplished in a second stage. Such a system is illustrated in Figure 3-11.

Use of aerobic digestion to reduce the amount of excess activated sludge to be further processed, while maintaining or improving its inherent processability, has been demonstrated at several plants. Metro Denver [16] is a particularly graphic example. Where there are existing anaerobic digestion facilities or all sludge streams must be stabilized prior to dewatering for final disposal, the process illustrated in Figure 3-12 might well be optimum.

In this system activated sludge is excluded from the anaerobic digester train. Blending of the two stabilized sludges just prior to dewatering eliminates problems associated with attempts to process mixed sludges.

3.4.6 Processing of Mixed Primary and Oxygen Activated Sludges

The oxygen activated sludge process is a staged co-current gas-liquid flow system with an improved capability for complete and simple contacting of pure oxygen with biomass. It is more favorable than conventional air processes for endogenous respiration and hence well bioflocculated sludge. Settling rate data for high purity oxygen and air biomasses are shown in Figure 3-13.

It would be expected that improved sludge characteristics afforded by pure oxygen would carry over into subsequent thickening and dewatering operations. Experience shows that this has occurred. Gravity thickening data follows in Table 3-16.

TABLE 3-16

GRAVITY THICKENING DATA FOR AIR AND
OXYGEN ACTIVATED SLUDGES [18]

Feed Sludge		Solids Loading	Underflow Conc.	
Type	Percent Solids	(lb/ft^2/day)	Percent Solids	Location
Oxygen EAS	1.7	10	4.8	Louisville[a]
Air EAS	0.9	20	1.4-2.8	Chicago
Oxygen mixed	2.3	--	5.6	Middlesex
Air mixed	1.1	20	3.3(4.4)[b]	Chicago

[a] Pilot plant data.

[b] Data in parentheses obtained with use of Picket Stirrer.

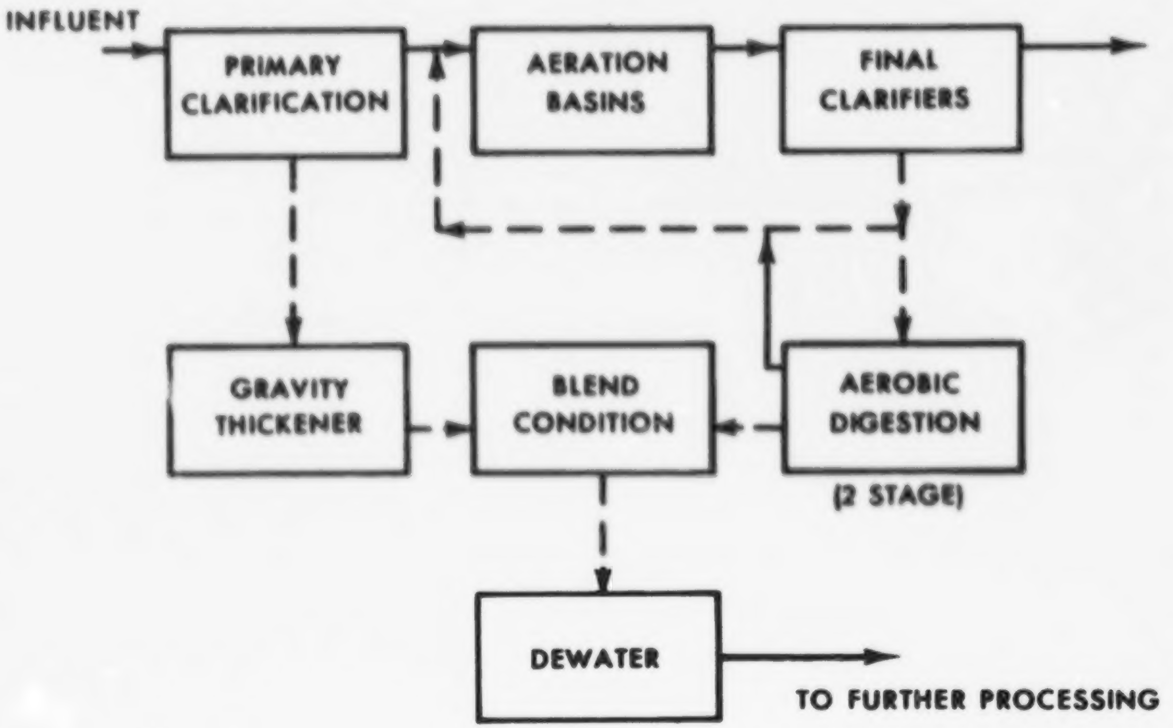

INFLUENT
PRIMARY CLARIFICATION
AERATION BASINS
FINAL CLARIFIERS
GRAVITY THICKENER
BLEND CONDITION
AEROBIC DIGESTION
(2 STAGE)
DEWATER
TO FURTHER PROCESSING

FIGURE 3-11. Use of aerobic digestion to reduce activated sludge solids in mixed sludge processing.

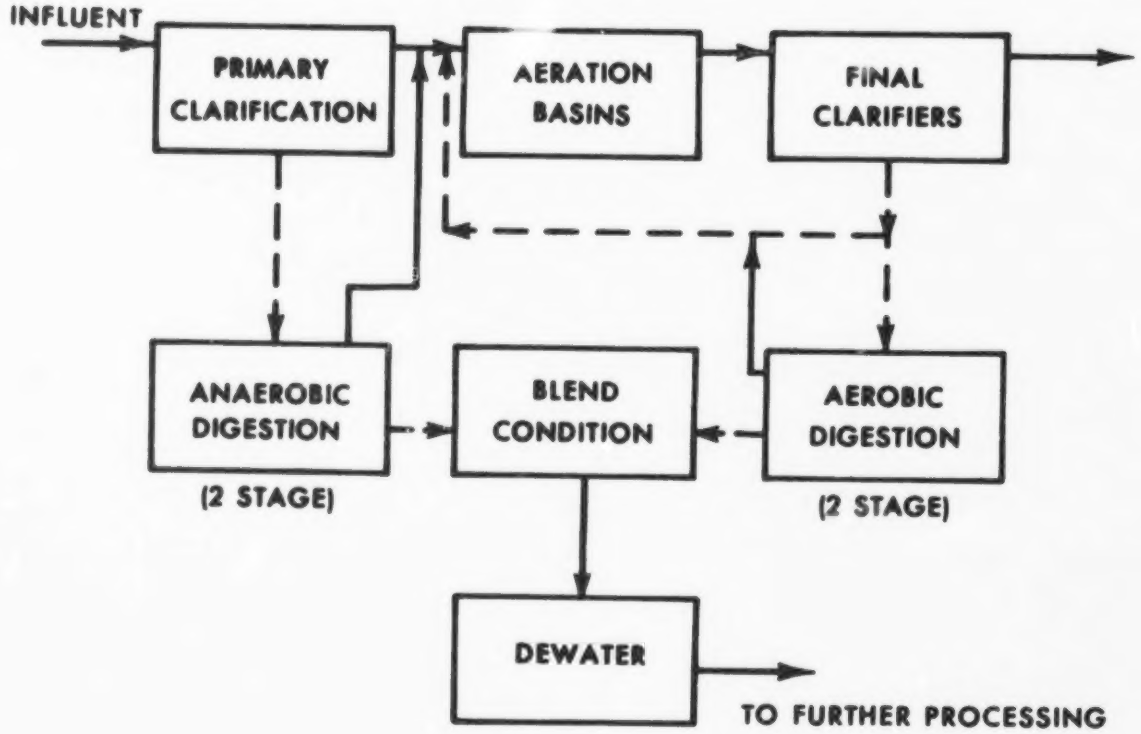

INFLUENT
PRIMARY CLARIFICATION
AERATION BASINS
FINAL CLARIFIERS
ANAEROBIC DIGESTION
(2 STAGE)
BLEND CONDITION
AEROBIC DIGESTION
(2 STAGE)
DEWATER
TO FURTHER PROCESSING

FIGURE 3-12. Improved use of anaerobic and aerobic digestion in mixed sludge processing.

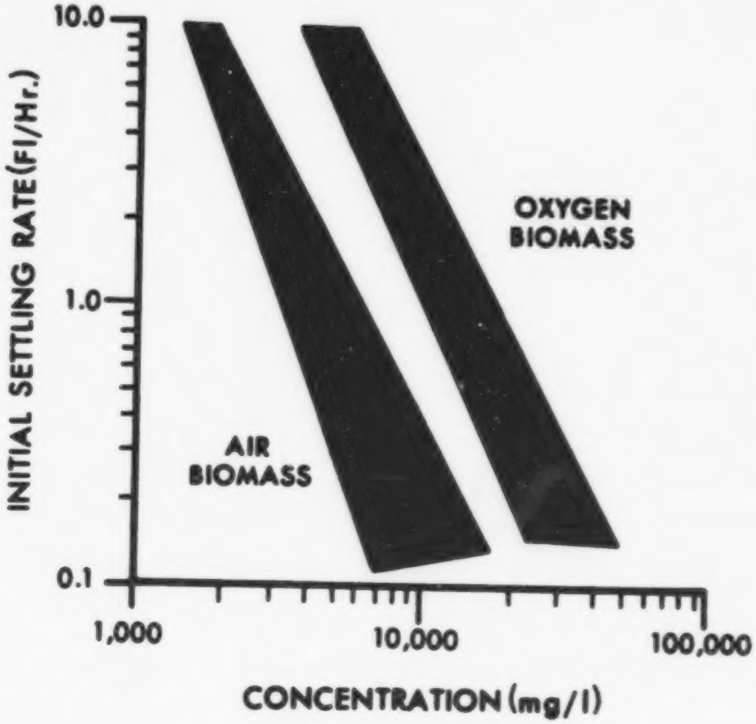


FIGURE 3-13. Settling characteristics for air and oxygen biomass (initial settling rate vs. concentration) [17].

While this data is admittedly on different sludges, the following data in Table 3-17 on air flotation thickening also indicates the improved processing properties for oxygen activated sludge.

TABLE 3-17

FLOTATION THICKENING DATA FOR AIR AND OXYGEN ACTIVATED SLUDGES [18]

Feed Sludge		Polymer (lb/ton)	Loading ($lb/ft^2/hr$)	Thickened Solids (%)
Type	Percent Solids			
Oxygen activated	1.7	2.9	6.4-10.2	6.6
Air activated	0.9	9.0	2.0-4.0	4.5

A typical comparison of dewatering information is shown in Table 3-18. This pilot data also indicates improved processability of oxygen activated sludge.

TABLE 3-18

CENTRIFUGATION DATA FOR OXYGEN AND CONVENTIONAL AERATION SLUDGES [18]

	Feed		Polymer (lb/ton)	Solids Capture (%)	Cake Solids (%)
Type Sludge	Percent Solids	Rate (gpm)			
Oxygen EAS	2.5	95	3	92	9
Air EAS	1.0	60	12.5	82	8.5

In the case of the Fairfax County, Virginia, Westgate Plant, the improved aspects of oxygen activated sludge have enabled use of the process flow diagram shown in Figure 3-14. This type system has shown significant economies. The Westgate Plant history is more fully described in Chapter 10.

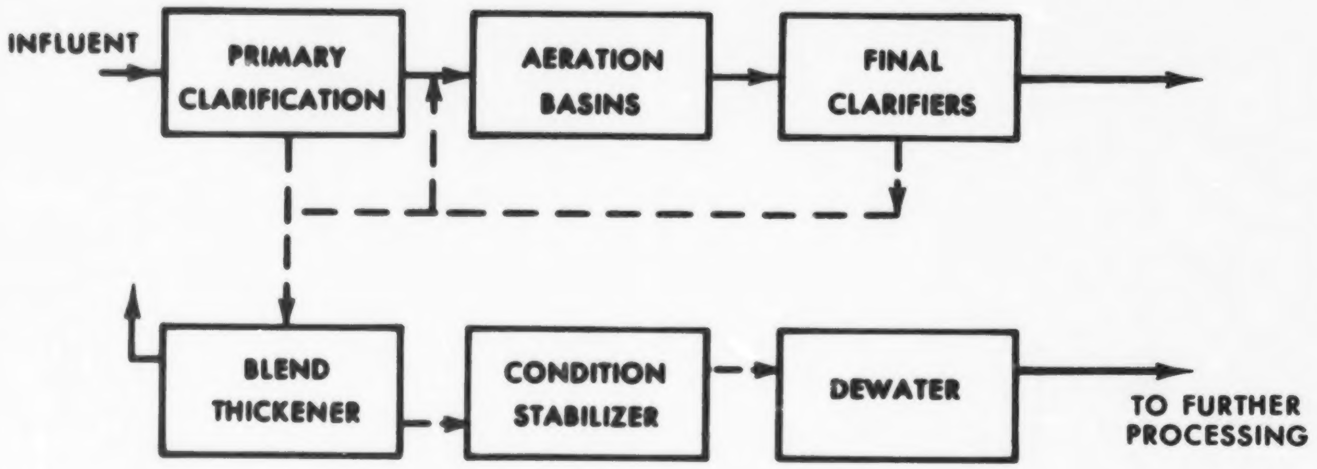

INFLUENT
PRIMARY CLARIFICATION
AERATION BASINS
FINAL CLARIFIERS
BLEND THICKENER
CONDITION STABILIZER
DEWATER
TO FURTHER PROCESSING

FIGURE 3-14. Oxygen activated sludge system at Westgate treatment plant.

3.4.7 Cake Release in Dewatering

A very important sludge processing characteristic which is sometimes overlooked, is the relative ease of release of filter cake from filter media. Failure to adequately consider this property has resulted in considerable difficulty and excess costs at several plants. While it is sometimes possible to alter this property of sludge by process changes, an important consideration is selection of the type of filter media.

Of the three types of rotary vacuum filters available, only the belt filter causes a severe problem with cake release. In contrast to the drum filter which features a scraper blade discharge from a blow-back section, and the coil filter which uses tines for release, the belt filter has essentially no release mechanism. Among plants which have experienced this problem are: Indianapolis; Baltimore; Richmond, California; Toronto; Washington, D.C.; and Columbus, Ohio. A direct comparison has occurred at Washington, D.C. Table 3-19 shows production rates and chemical costs associated with use of drum filters versus belt filters at Washington, D.C.

TABLE 3-19

PERFORMANCE DATA FOR DRUM FILTERS AND BELT FILTERS AT WASHINGTON, D.C.

Type Filter	Yield (lb/ft^2/hr)	Flocculation Cost ($/ton)*
Drum	3.5	5.65
Belt	3.5	8.80
Drum	4.0	3.95
Belt	4.0	7.75

*Based on 1971 cost figures.

As can be seen, the belt filters consumed substantially more ferric chloride to promote release than the drum filters with the identical sludge.

3.4.8 Phosphorus Removal Process Sludges

As noted earlier, application of metallic salts at various points in biological treatment plants for phosphorus removal creates new and additional sludges. Thorough plant scale data on the processing characteristics of such sludges are not widespread. Plant scale work at the North Toronto, Ontario plant using ferric chloride has produced some results of note. The ferric chloride was applied at the end of the aeration basins.

TABLE 3-20

DEWATERING DATA ON NORTH TORONTO FERRIC CHLORIDE–ACTIVATED SLUDGE [19]

Chemical Addition		Mixed Sludge Percent Solids	Cond. (lb/ton)		Filter Yield ($lb/ft^2/hr$)	Percent Cake Solids
Metal Salt	Dose (mg/l)		Ferric Chloride	Lime		
Ferric Chloride	25-35	8	104	200	3.3	21

As seen in Table 3-20 a reasonable processing rate, cake solids concentration, and chemical conditioning were achieved.

At the Newmarket, Ontario activated sludge plant, lime was used in the primaries with the sludge generation and processing results shown in Table 3-21. Indicated results with the mixture of limed primary and activated sludges at Newmarket were excellent. Good solids capture and concentration were realized at low cost. The amount of sludge mass for disposal tripled.

TABLE 3-21

DEWATERING DATA ON NEWMARKET MIXED ORGANIC/LIME SLUDGE [20]

Chemical Addition		Mixed Sludge Percent Solids	Solids Produced (tons/M.G.)	Centrifugation		
Metal Salt	Dose (mg/l)			Polymer (lb/ton)	Percent Cake Solids	Solids Capture (%)
None	–	3.5	0.85	–	–	–
Lime	200	10	2.45	<1	31	97

A direct comparison of sludge handling for the case of lime addition to the primary process and the case of alum addition to the aeration process was made at the Windsor Little River conventional activated sludge plant and is summarized in Table 3-22. First, note that the untreated system chemical conditioning costs were abnormally high at $16/ton, even though a good feed sludge solids concentration to the dewatering step was realized. Lime usage gave 50 percent additional sludge, but dewatering costs per ton were lowered. Alum use caused a lower filter yield and higher dewatering costs.

TABLE 3-22

SLUDGE PROCESSING DATA FOR PHOSPHORUS REMOVAL ALTERNATIVES AT THE LITTLE RIVER TREATMENT PLANT [21]

Chemical Addition		Mixed Sludge Percent Solids	Solids (tons/M.G.)	Filter Yield (lb/ft²/hr)	Conditioning Cost ($/ton)*
Metal Salt	Dose (mg/l)				
None	–	6.2	0.8	5.2	16
Lime	125	11.6	1.2	7.2	11
Alum	150	5.7	1.2	4.6	18

*Based on 1972 cost figures.

3.5 Additional Reading

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U.S. Patent No. 3,763,039 October 2, 1973, "Method of Treating Sewage Using High Polymer Ratio Flocculation Agent Biologically Produced in Situ," Inventor George E. Wilson, Sterling, Illinois, assigning Houdaille Industries Incorporated, Buffalo, New York.

Wallen, L. L., Davis, E. N., "Biopolymers of Activated Sludge," *Environ. Sci. Technol.* 6(2), 161 (Feb. 1972).

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17. U.S. Environmental Protection Agency, Technology Transfer Seminar Publication, "Oxygen Activated Sludge Wastewater Treatment Systems, Design Criteria and Operating Experience" (Aug. 1973).

18. EPA-TTP Design Seminar for Sludge Processing and Disposal, Atlanta, Georgia, Dec. 11-12, 1973.

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21. Van Fleet, G. L., Barr, J. R., and Harris, A. J., Ontario Ministry of the Environment, "Treatment and Disposal of Chemical Phosphate Sludges in Ontario." WPCF Meeting, Atlanta, Georgia, Oct. 1972.

CHAPTER 4

SLUDGE THICKENING (BLENDING)

4.1 Functions, Methods, and Occurrences

Thickening increases the solids content of a sludge stream by partial removal of the liquid portion. During sludge handling this process frequently occurs more than once. The purpose [illegible]duce the sludge volume to be stabilized, dewatered, or hauled away as a liquid to final disposal. Figure 4-1 shows the effect of thickening on sludge volume.

Most benefits of reduced sludge volume on subsequent treatment processes, such as anaerobic digestion, are illustrated in Figure 4-2. The results in Figure 4-2 were derived from a two-stage trickling filter plant at Beaumont, Texas. Effect of increased sludge solids content on dewatering is further documented in Chapter 7.

Figure 4-3 shows the importance of having a high solids sludge going to a mechanical dewatering device. These results of McCarty's [1] are from a large number of plants processing different sludges. As noted earlier, thickening occurs as an extra benefit in unit processes provided essentially for another purpose. Individual unit processes that afford sludge thickening are summarized in Table 4-1.

TABLE 4-1

OCCURRENCE OF THICKENING IN WASTEWATER TREATMENT PROCESSES

Unit Process	Principal Functions
Primary Sedimentation	Clarification-liquid, sometimes blend and/or thicken sludge
Elutriation Basin	Wash and thicken sludge
Secondary Sedimentation	Clarification-liquid, partially thicken sludge
Gravity Thickener	Thicken and blend sludge
Dissolved Air Flotation Thickener	Thicken sludge
Centrifuge Thickener	Thicken sludge

Frequently, inadequate attention has been given to sludge processing considerations in the design of sedimentation units. Dick and Ewing [3], Dick [4,5], and Mancini [6] have graphically discussed the prevalent failure to consider both the clarification and thickening aspects in design of sedimentation units.

The relative usage of various methods of sludge thickening in sludge treatment are delineated in Table 4-2.

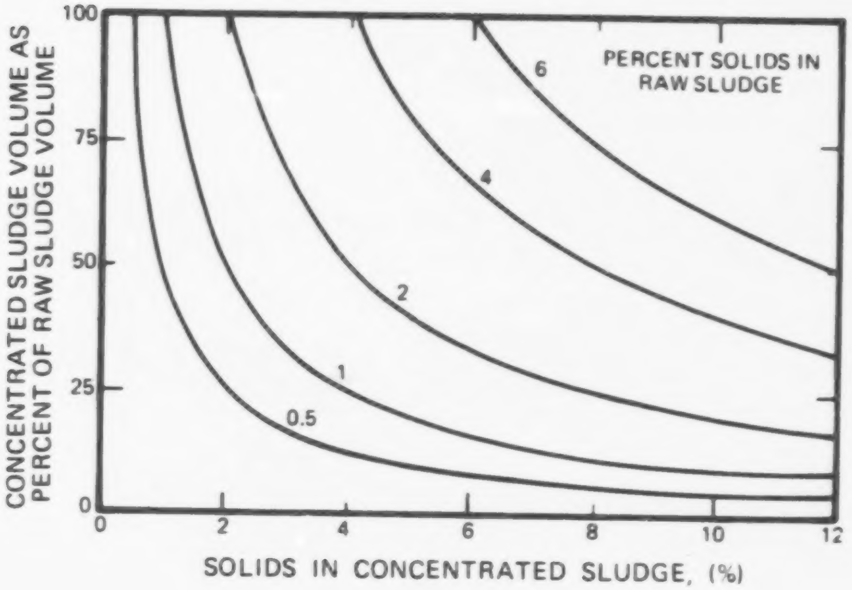


FIGURE 4-1. Effect of increasing sludge solids on the final sludge volume [1].

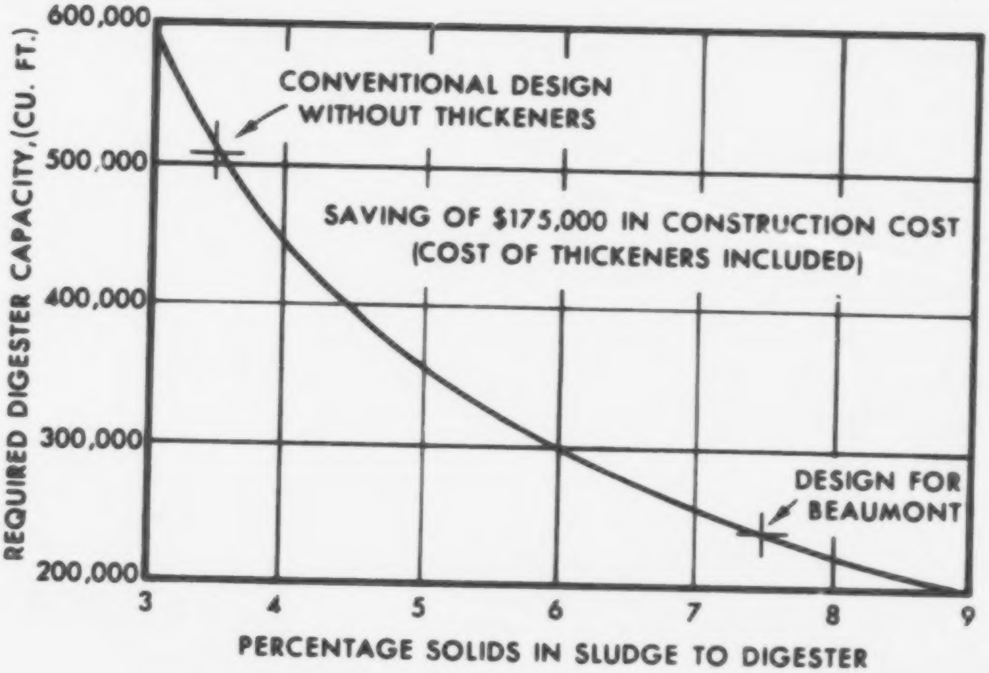


FIGURE 4-2. Effect of thickening on required digestion capacity [2].

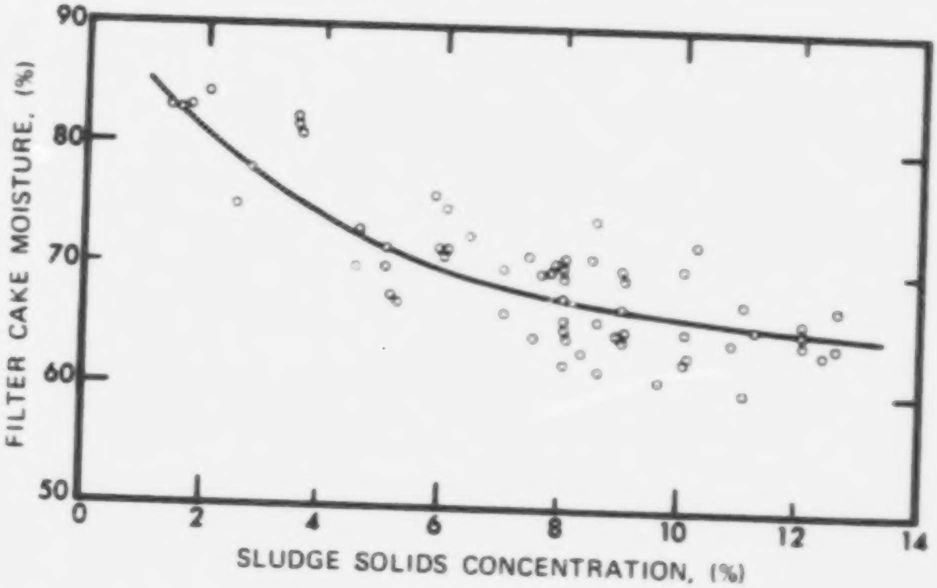


FIGURE 4-3. Effect of feed solids on performance of a rotary vacuum filter [1].

TABLE 4-2

OCCURRENCE OF THICKENING METHODS IN SLUDGE TREATMENT

Method	Type Sludge	Other Pertinent Considerations	Frequency of Usage and Relative Success
Gravity	Raw Primary	Separate Air EAS Thickening	Increasing - excellent results
Gravity	Digested Primary	Separate EAS Thickening	Infrequent now, but feasible
Gravity	Raw Primary and EAS	Recirculation of Air EAS to Primaries	Decreasing usually poor
Gravity	Raw Primary and EAS	Direct Mixed Sludge Thickening	Some new installations marginal results
Gravity	Air EAS	Separate Primary Sludge Thickening	Essentially never used poor results
Gravity (Elutriation)	Digested Primary and EAS Mixture	Secondary Digesters or Elutriation	Many plants built requires flocculants
Dissolved Air Flotation	Raw Primary and EAS	Direct Mixed Sludge Thickening	Rarely used
Dissolved Air Flotation	Air EAS	Separate Gravity Thickening of Primary	Increasing - good results
Solid Bowl Conveyor Type Centrifuge	Air EAS	Separate Gravity Thickening of Primary	Some limited use - solid capture problem
Disc Type Centrifuge	Air EAS	Separate Gravity Thickening of Primary	Some limited use - data now being accumulated

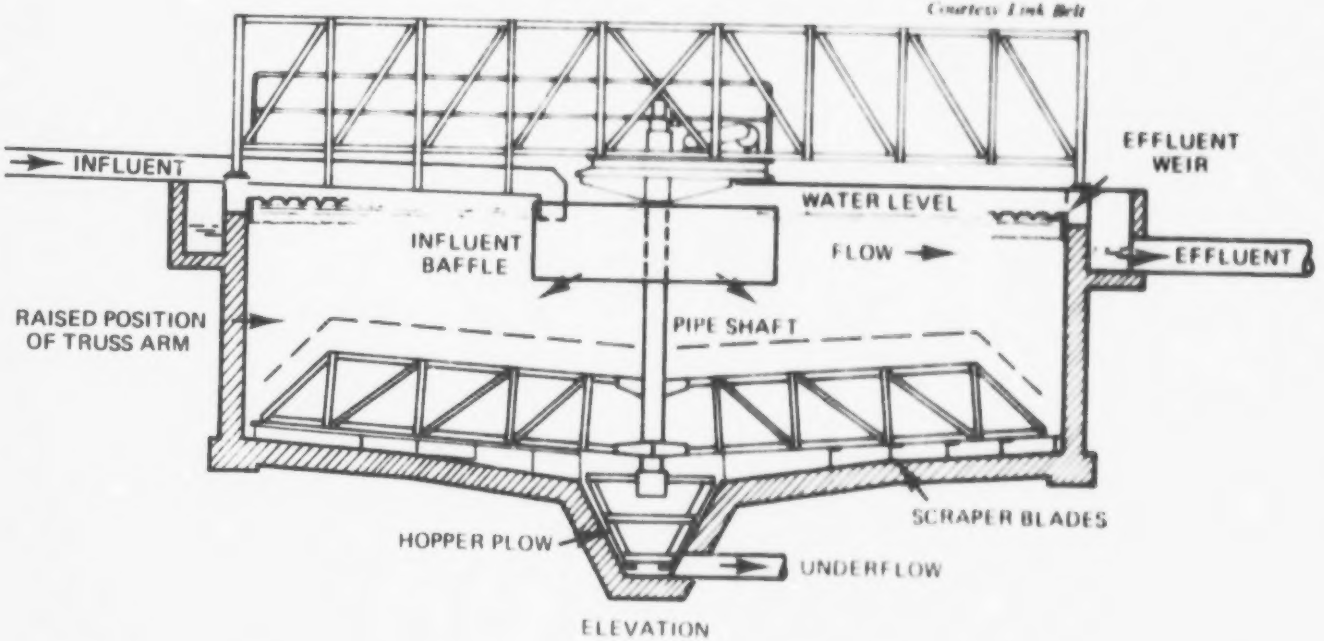
Gravity thickening of raw or digested primary sludge is almost always an efficient and economical process. Anaerobically digested primary sludge is normally thickened by gravity in the secondary digester. As discussed in detail in Chapter 10, the use of primary basins to capture and thicken both wastewater influent and recirculated excess activated sludge solids has been found to be poor practice. The EAS solids do not resettle well in the primary, and this results in the production of more EAS due to an increased load on the aeration system. Further, relatively poor thickening results when the primary basins happen to capture some of the EAS solids. Because the gravity thickeners process mixtures of raw primary and air system excess activated sludges inefficiently and only provide marginal solids capture, they are infrequently used at this time. The relative degree of difficulty encountered has been correlated to the particular type of activated sludge system used and the settling properties of the excess activated sludge. Oxygen activated sludge, however, has been shown in limited plant scale testing to be more amenable to mixed sludge gravity thickening. The efficiency of gravity thickening of mixed liquor in secondary sedimentation basins, as shown by current operating data is poor. Therefore a further need for thickening of EAS exists. This is true for EAS from air and pure oxygen systems. The city of Grand Rapids, Michigan performed a plant study which included an unsuccessful attempt to use gravity thickeners for straight activated sludge [7].

Elutriation had been regarded generally as a means for reducing the demand for acidic chemical conditioner by washing out the fine solids and alkalinity. It is now normally used also to thicken combined primary and secondary sludges after anaerobic digestion. Such thickening is accomplished with lower volumes of wash water and polyelectrolyte addition. In fact, elutriation is now well recognized as a postdigestion gravity thickening process.

Use of dissolved air flotation for thickening of EAS has increased because it gives reliable and effective results. Centrifugal thickening of EAS has not yet been widely applied. Recent improvements, however, in the design of solid bowl conveyor units and disc units may alter this situation.

4.2 The Gravity Thickener

A conventional gravity thickener is shown in Figure 4-4. The thickener normally consists of two truss-type steel scraper arms mounted on a hollow pipe shaft keyed to a motorized hoist mechanism. A truss-type bridge is fastened to the tank walls, or, in some cases, to steel or concrete columns, spans the tank, and supports the entire mechanism. The thickener resembles a conventional circular clarifier with the exception of having a greater bottom slope. Sludge enters at the middle of the thickener and the solids settle into a sludge blanket at the bottom. The concentrated sludge is very gently agitated by the moving rake which dislodges gas bubbles, prevents bridging of the sludge solids, and keeps the sludge moving toward the center well from which it is removed. Supernatant liquor passes over an effluent weir around the circumference of the thickener. McCarty [1] found that in the operation of gravity thickeners it is desirable to keep a sufficiently high flow of fresh liquid entering the concentrator to prevent septic conditions and resulting odors from developing. This can be done by specifying an overflow rate of between 600 and 800 gpd/sq ft. To achieve hydraulic loadings in this range, secondary effluent is normally blended with the sludge feed to the thickener. As mentioned previously, chlorine also can be used for septicity prevention. Another design factor is the sludge volume ratio (SVR) which was defined by Torpey [9] to be the volume of the sludge blanket divided by the daily volume of sludge pumped from the thickener. This relationship has the units of days and is used as a relative measure of the average retention time of solids in the thickener. A long SVR is desirable for maximum concentration but


INFLUENT
EFFLUENT WEIR
WATER LEVEL
EFFLUENT
INFLUENT BAFFLE
FLOW
RAISED POSITION OF TRUSS ARM
PIPE SHAFT
SCRAPER BLADES
HOPPER PLOW
UNDERFLOW
ELEVATION

Courtesy Link Belt

FIGURE 4-4. Gravity thickener [8]

may lead to excessive biological decomposition. Values for SVR are normally maintained between 0.5 and 2 days, with the lower values being used during warmer weather. Solids loadings based on lb/hr/ft^2 vary with the type of sludge. Most continuous thickeners today are circular and designed with a side water depth of approximately 10 feet. While sludge blanket depth is an important parameter, it has been reported that underflow solids concentrations are independent of sludge blanket depths greater than 3 feet. Increased sludge detention time in the thickener will result in increased underflow solids concentration. A detention period of 24 hours has been suggested as the time required to achieve maximum compaction. Sludge blanket depth and detention time are closely interrelated. The sludge blanket depth may be varied with fluctuation in solids production to achieve good compaction. During peak conditions the detention time may have to be shortened to keep the sludge blanket depth sufficiently below the overflow weirs to prevent excessive solids carry-over.

4.2.1 Performance Experiences

Typical solids loadings as well as thickener output concentration for various sludge types were summarized by Newton [10] and are shown in Table 4-3.

TABLE 4-3

GRAVITY THICKENER SURFACE LOADINGS AND OPERATIONAL RESULTS [10]

Type of Sludge	Solids—Surface Loading (lb/day/ft^2)	Thickened Sludge Solids Concentration (%)
Separate sludges		
Primary	20-30	8-10
Modified activated	15-25	7-8.5
Activated	5-6	2.5-3
Trickling filter	8-10	7-9
Combined sludges		
Primary and modified activated	20-25	8-12
Primary and activated	6-10	5-8
Primary and trickling filter	10-12	7-9

This data should be viewed carefully since:

- No data on quality of thickener overflows is presented and both loading and the degree of thickening accomplished would influence this.
- Achievement of only 2 to 3 percent solids in the gravity thickening of activated sludge further shows the inadequacy of this method for such sludges.
- Data on combined sludges is reasonable, except that few plants attain a 5 to 8 percent thickened sludge solids concentration with a mixture of primary and activated sludges.

4.2.2 Theory of Gravity Thickening and Design Procedures

Mancini [6], and later Dick and Ewing [3], Dick and Young [11], Dick [4,5], and Edde, Eckenfelder, and Wesley [12] called attention to the inadequacy of the Coe and Clevenger [13], Kynch [14], and Talmage and Fitch [15] procedures when significant amounts of activated sludge are involved in gravity thickening operations. It is frequently necessary, in industrial and domestic waste treatment, to provide clarification and thickening facilities for flocculent slurries which have hindered settling characteristics. These slurries form a distinct solids-liquid interface when allowed to settle on a batch basis. A plot of the elevation of the solids-liquid interface as a function of time provides a characteristic hindered settling curve as shown in Figure 4-5.

A method of designing continuous flow clarification and thickening facilities for suspensions which have hindered settling characteristics has been proposed by Kynch [14] and modified by Talmage and Fitch [15]. A batch settling test is performed and a settling curve is analyzed to determine the surface area required for clarification and the surface area required to thicken the slurry to a desired underflow solids concentration. Details of the classical method of design of gravity thickeners based on the analysis of batch settling tests are fully described in the various references previously cited. It is sufficient here to give results and conclusions by Mancini [6] and Dick et al. [3,4,5,11] from studies on various types of sludges. The conventional thickening concepts as outlined have been generally accepted, without qualification, as being applicable to solids concentration problems in sanitary engineering.

Kynch [14], however, presumed all particles were of the same size and shape, and that they were uniformly distributed in any horizontal plane. He did not discuss flocculent particles or the applicability of the theory to compressible materials.

In contrast to the suspension considered by Kynch [14], activated sludge is comprised of nonrigid, flocculent particles. The floc combine into aggregate particles whose size is not necessarily uniform nor spherical, and is subject to change as dictated by the local shear rate. Furthermore, and possibly of most importance, it seems likely that a network of

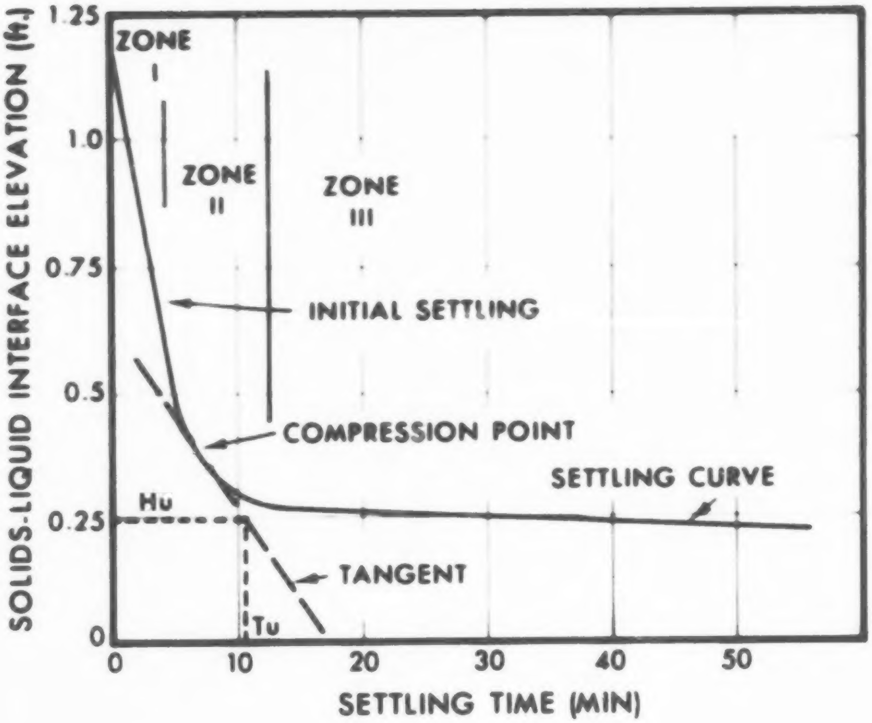


FIGURE 4-5. Characteristic settling curve for slurry with hindered settling characteristics.

aggregate particles could form a structured suspension exhibiting a yield strength. The existence of such a structure would permit interparticle forces. Yet Kynch's basic assumption is equivalent to an assumption that the only forces on particles are of hydraulic origin.

Mancini [6] studied the gravity settling properties of several activated sludges in detail. His major conclusions were:

- The Kynch [14] theory is valid for ideal suspensions. An example is the behavior of sand suspensions.

- The settling behavior of activated sludge cannot be predicted by the Kynch theory. Hence, the prevailing theories of thickening are not strictly applicable to the design of final settling tanks and gravity thickeners for activated sludge.

- In addition to being dependent upon concentration, the rate of subsidence of activated sludge is dependent upon sludge depth and the mixing of underlying layers. This is true at concentrations less than that of the compression point.

- Analysis of the relationship between the initial depth of an activated sludge suspension and its initial settling velocity provides quantitative measures of the extent of the deviation from ideal behavior (the retardation factor) and the ultimate settling velocity of the suspension.

- Within the concentration ranges investigated, the retardation factor of activated sludges varied exponentially with sludge concentration.

- The retardation factor can be related to the general nature of activated sludge. At comparable concentrations, low retardation factors are associated with sludges of good settleability, while bulking sludges have high retardation factors.

- Settling tests for determining the solids handling capacity of activated sludge must be conducted with depths and mixing conditions comparable to plant operating conditions.

Plant experiences generally agree with the conclusions presented. Definitive work on analysis of design procedures for gravity thickening has been continued by Dick [5]. These procedures will also still be useful in considering gravity thickening of mixtures of primary sludge and excess oxygen activated sludge.

4.2.3 Gravity Thickening of Oxygen Activated Sludges

Successful gravity thickening of a mixture of raw primary and excess oxygen activated sludge is practiced at the Westgate treatment plant in Fairfax County, Virginia. This plant's operating experiences are discussed in detail in Chapter 10. Pilot data on separate gravity thickening of oxygen activated sludge shows some promise as illustrated in Table 4-4.

TABLE 4-4

GRAVITY THICKENING DATA FOR EXCESS OXYGEN ACTIVATED SLUDGE

Location	Feed Solids Concentration (%)	Solids Loading (lb/ft^2/day)	Thickened Solids Concentration (%)
Batavia	1.7 - 3.0	24	4.5 - 6.0
Tonawanda	3.0 - 3.5	34	7.0 - 8.0
New Orleans	2.5 - 3.0	15	6.0
Louisville	1.8 - 3.0	12 - 23	4.5 - 5.0
Philadelphia	1.5 - 2.0	11 - 15	4.0 - 5.0
Middlesex	1.5 - 2.0	7 - 9	4.0 - 4.5

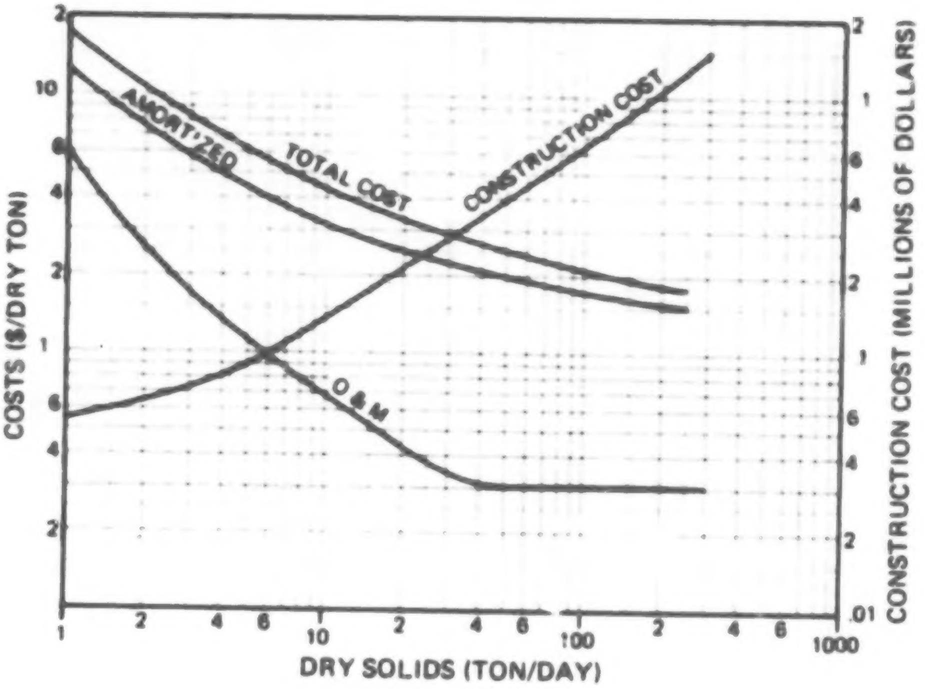
The high thickened sludge solids content and the solids loading rates achieved, indicate that gravity thickening with this type sludge may well be applicable. Further plant scale data if possible and delineation of overflow analyses should precede routine use of this method.

4.2.4 Capital, Operation, and Maintenance Costs for Gravity Thickening

Gravity thickening costs were reported by Stanley Consultants, Inc. [16], in November, 1972, and these are presented in Figure 4-6.

4.3 Air Flotation Thickening

In general, air flotation thickening can be employed whenever particles tend to float rather than sink. These procedures are also applied if the materials have a long subsidence period and resist compaction for thickening by gravity.



NOTES:

1. Minneapolis. Mar., 1972. ENR Construction Cost Index of 1827.
2. Amortization at 7% for 20 years.
3. Labor rate of $6.25 per hour.
4. No chemicals.
5. Influent sludge with a solids content of 0.8%.
6. Source: Envirotech and Stanley Consultants.

FIGURE 4-6. Costs of gravity thickening.

4.3.1 Occurrence, Methods, and Process Theory

Flotation thickening units as reviewed by Culp [17] are becoming increasingly popular for handling excess activated sludges since they have the advantage over gravity thickening tanks of offering higher solids concentrations and a lower initial equipment cost. Flotation, which uses rising gas bubbles to increase the buoyance of solid particles, is accomplished in four ways:

1. Dispersed air flotation occurs when bubbles are generated by introducing air through a revolving impeller or porous media.

2. Dissolved air-pressure flotation occurs when air is put into solution under elevated pressures and later released at atmospheric pressure.

3. Dissolved air-vacuum flotation occurs when a vacuum is applied to wastewater aerated at atmospheric pressure.

4. Biological flotation occurs when the gases formed by natural biological activity are used to float solids.

Dissolved air-pressure flotation will be discussed in detail in this section. Of the preceding techniques, it enjoys by far the widest usage for thickening sludge in the United States.

The objective of flotation-thickening is to cause the solids to separate from the water in an upward direction by attaching minute air bubbles to particles of suspended solids. The solid particles with attached bubbles have a specific gravity lower than water and tend to float. The bubbles formed must have a small diameter and this is accomplished by releasing air from a solution that has been pressurized at 40 to 80 psi. Since the solubility of air increases with pressure, substantial quantities of air can be dissolved. In modern flotation practice, two general approaches to pressurization are:

- Air charging and pressurization of dilution water (frequently recycled clarified effluent) with subsequent addition to the feed sludge.

- Air charging and pressurization of the combined dilution liquid and feed sludge.

Release of the pressurized flow into a chamber at near atmospheric pressures decreases solubility of air, and the excess comes out of solution to form minute air bubbles (average diameter 80 microns). These attach themselves to and are enmeshed in particles of flocculated sludge.

There are several manufacturers of dissolved air flotation (DAF) systems used for wastewater sludge thickening. Three widely used systems in the United States are those manufactured by (1) Komline-Sanderson, (2) Rexnord, and (3) Envirotech. Figure 4-7 shows a typical unit and a typical system installation is illustrated in Figure 4-8.

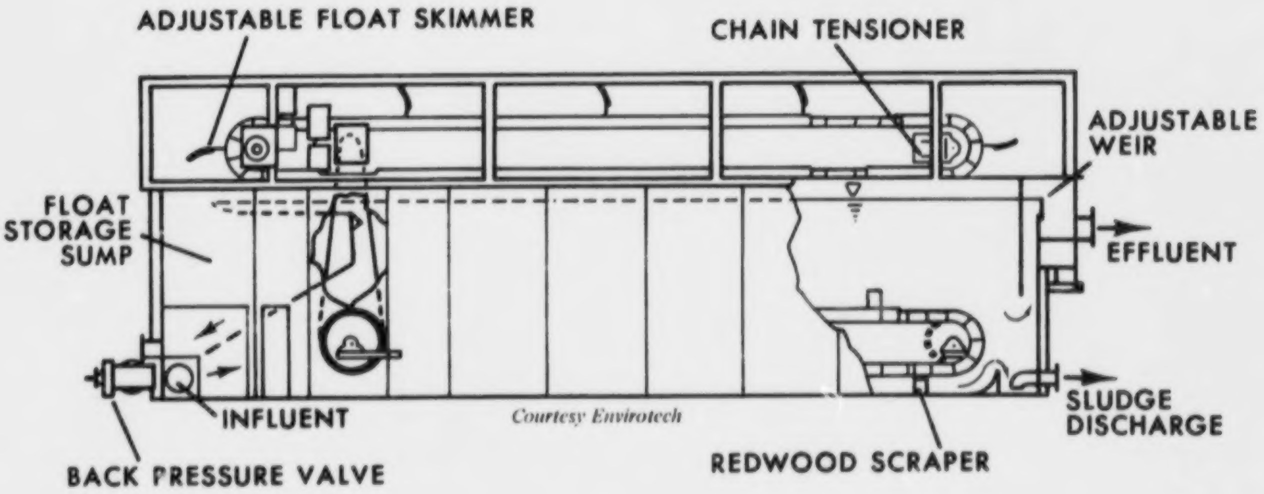
Figure 4-8 is based on the Komline-Sanderson system and serves to illustrate the basic process considerations. In their system a portion of the flotation unit's effluent, or similar plant process stream, is pumped to a retention tank at 60 to 70 psig. Air is fed into the pump discharge line at a controlled rate and mixed by the action of an eductor driven by the reaeration pump. The flow through the recycle system is metered and controlled by a valve located immediately before the mixing of the recycle stream with the sludge feed. Effluent recycle ratios can range from 30 to 150 percent of the influent flow. The recycle flow and sludge feed are mixed in a chamber at the unit inlet. If flotation aids are employed, introduction is normally in this mixing chamber. The sludge particles are floated to the sludge blanket and the clarified effluent is discharged under a baffle and over an adjustable weir which controls the depth of penetration of the skimming blades. The thickened sludge is removed by a variable speed skimming mechanism. In practice, bottom sludge collectors are also furnished for removal of any settled sludge or grit that may accumulate.

Sludge thickening occurs in the sludge blanket, which is normally 8 inches to 24 inches thick. The buoyant sludge and air bubbles force the surface of the blanket above the water level, inducing drainage of water from the sludge particles. Detention time in the flotation zone is not critical, providing the particle rise rate is sufficient and that horizontal velocity in the unit does not produce scouring of the sludge blanket.

Design and mode of operation of the pressurization tanks differ among the Komline-Sanderson, Rexnord, and Envirotech systems. The Envirotech and Rexnord systems inject the liquid into the air to get the desired air dissolution while Komline-Sanderson injects the air into the liquid. Both approaches are satisfactory and the operating pressure of the pressurization tank is a more important variable.

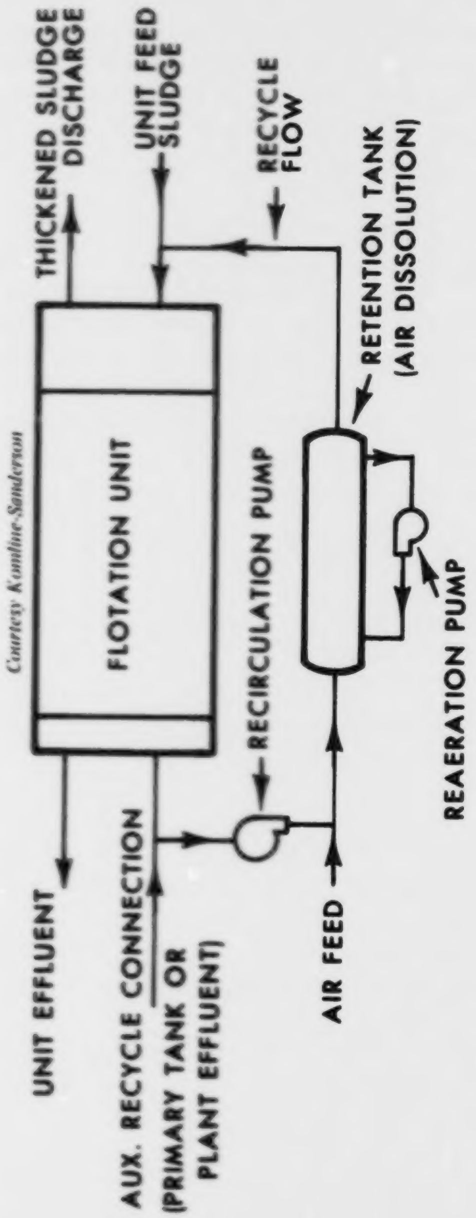
The primary variables for flotation thickening are:

- Pressure
- Recycle Ratio
- Feed Solids Concentration
- Detention Period
- Air-to-Solids Ratio
- Type and Quality of Sludge


ADJUSTABLE FLOAT SKIMMER
CHAIN TENSIONER
ADJUSTABLE WEIR
FLOAT STORAGE SUMP
EFFLUENT
SLUDGE DISCHARGE
INFLUENT
BACK PRESSURE VALVE
REDWOOD SCRAPER

Courtesy Envirotech

FIGURE 4-7. Dissolved air flotation unit.


THICKENED SLUDGE DISCHARGE
UNIT FEED SLUDGE
RECYCLE FLOW
RETENTION TANK (AIR DISSOLUTION)
FLOTATION UNIT
RECIRCULATION PUMP
REAERATION PUMP
PUMP
UNIT EFFLUENT
AUX. RECYCLE CONNECTION (PRIMARY TANK OR PLANT EFFLUENT)
AIR FEED

Courtesy Komline-Sanderson

FIGURE 4-8. Dissolved air flotation system.

- Solids and Hydraulic Loading Rates
- Use of Chemical Aids

Air pressure used in flotation is important because it determines air saturation or size of the air bubbles formed. It influences the degree of solids concentration and the subnatant (separated water) quality. In general, either increased pressure or air flow produces greater float (solids) concentrations and a lower effluent suspended solids concentration. There is an upper limit, however, too much air breaks up fragile floc. The Komline-Sanderson and Envirotech systems operate at 60 to 70 psig while the Rexnord operates at 40 to 50 psig.

Operating pressure of the pressurization tank also has an effect on the size of the bubbles released in the thickener. If the bubbles are too large, there will not be enough surface area for attachment to the sludge to be floated. Generally, bubble sizes less than 100 microns in diameter, and preferably around 50 microns, are believed most desirable for effective flotation of sludge. Although bubbles below 100 microns can be seen with the naked eye, the only practical way to establish the size is by rise rate experiments. Such experiments have been published [14] and the results indicate that a faster rise rate occurs when air is released from a 40 psi pressure than from an 80 psi pressure, as shown in Figure 4-9. This indicates that larger bubbles are formed when air is released from lower pressures.

Recycle ratio and feed solids concentration are interrelated. Additional recycle of clarified effluent does two things:

1. It allows a larger quantity of air to be dissolved because there is more liquid.
2. It dilutes the feed sludge.

Dilution reduces the effect of particle interference on the rate of separation. At the Chicago Sanitary District, 40 percent recycle proved to be optimum. Concentration of sludge increases and the effluent suspended solids decrease as the sludge blanket detention period increases. In plant tests, solids concentration increased rapidly with time up to 3 hours. Beyond 3 hours, however, no additional thickening was observed.

The air-to-solids ratio is also an important parameter because it influences the sludge rise rate. Figure 4-10 presents data collected for several activated sludges resulting from the treatment of various type wastes.

The air-to-solids ratio needed for a particular application is a function primarily of the sludge's characteristics such as SVI. The most common ratio used for design of an excess activated sludge thickener is 0.02. Figure 4-10 indicates that for a sludge which will easily flocculate and settle (SVI below 100), an air-to-solids ratio of 0.03 should be sufficient to achieve a thickened solids concentration of 4 percent.

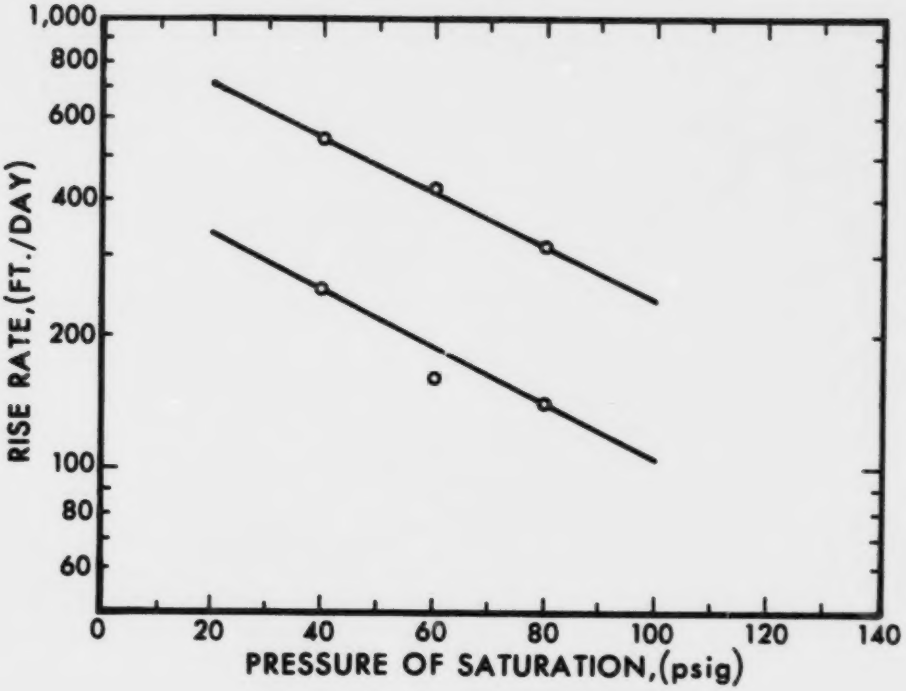


FIGURE 4-9. Influence of pressure of saturation on rise rate [14].

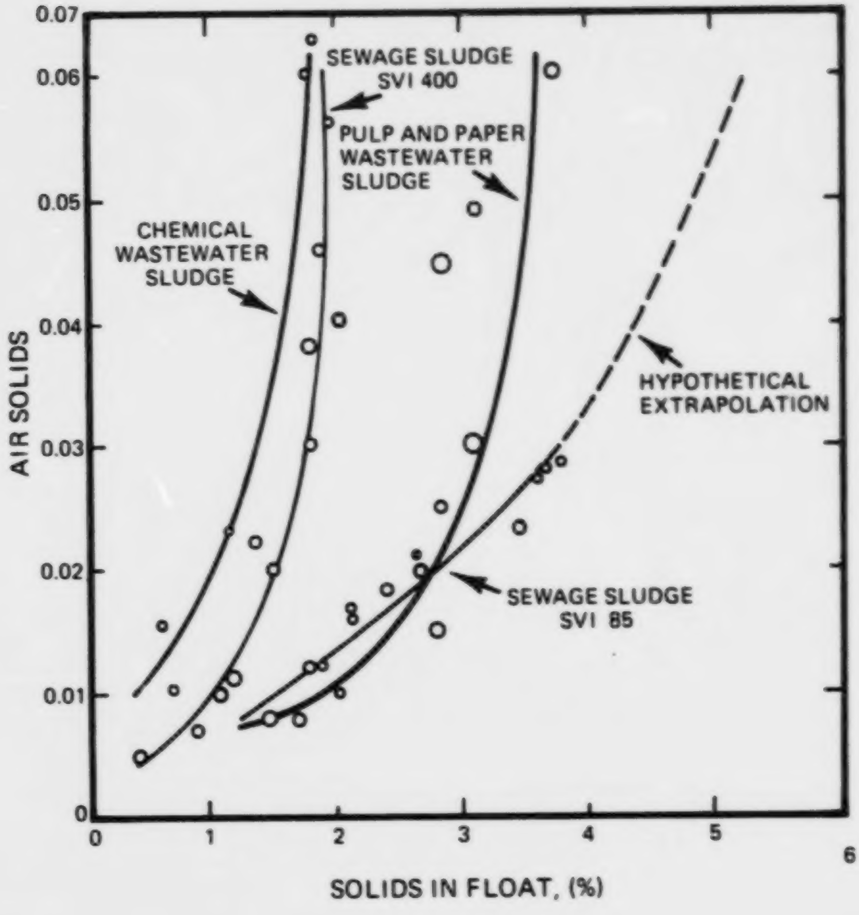


FIGURE 4-10. Influence of air-to-solids ratio on float solids content [14].

4.3.2 Operational Results

Like sedimentation, the type and quality of sludge to be floated affects unit performance. Flotation thickening is most applicable to activated sludges, as stated before, but higher float concentrations can be achieved with combined primary and activated sludge. A high SVI represents a bulky sludge and results in poor thickener performance.

Loading rate affects performance and Table 4-5 summarizes loadings employed and results experienced at several operating installations. Solids loading is the design parameter which determines the surface area of the unit. Generally, the higher the solids loading, the lower the solids in the thickened sludge.

Many different chemicals have been used in wastewater air flotation systems. Flocculating chemicals agglomerate solids into stable floc that promote an increase in the terminal velocity and facilitate capture of gas bubbles. Overall effect is to increase allowable solids loadings, percentage of floated solids, and improve the clarity of the subnatant. Cationic polyelectrolytes have been the most successful chemicals used in wastewater sludge thickening. Chicago, for example, found a polymer dosage of 12 lb/ton very helpful in the flotation thickening of an activated sludge with poor settling characteristics. Thickening of a chemically conditioned sludge with a SVI of 104 to 107 was comparable to that obtained with untreated activated sludge having a SVI of 84. Even higher loading rates (0.72 $lb/hr/ft^2$ vs. 0.58 $lb/hr/ft^2$) were achieved with the conditioned sludge. Experience has shown that the introduction of the polymer solution into the line just as the bubbles are being formed and mixed with the sludge produces the best results. The decision to use or not use polymers or other chemicals must be based on an optimization of capital costs versus the added operating costs from chemical usage.

The pronounced effect of polymer usage in flotation thickening of activated sludge is demonstrated in Table 4-5. In three of the four locations not using flotation aids, float cake solids were lower than a generally preferred minimum of 4 percent and the system loadings were also lower. Nearly all installations now use flotation aids.

4.3.3 Advantages and Disadvantages of DAF Thickeners

Commonly cited considerations in selecting DAF thickening over that of gravity thickening for processing activated sludge are:

- Reliability
- A thicker product
- A higher solids loading
- Lower capital cost

TABLE 4-5

OPERATING DATA FOR PLANT SCALE DAF UNITS [17]

Location	Feed	Influent ss mg/l	Subnatant ss mg/l	% Removal ss	Float % Solids	Loading lb/hr/ft^2	Flow gpm/ft^2	Remarks
Bernardsville, N.J.	M.L.[a]	3,600	200	94.5	3.8	2.16	1.2	Standard[c]
Bernardsville, N.J.	R.S.[b]	17,000	196	98.8	4.3	4.25	0.5	Standard
Abington, Pa.	R.S.	5,000	188	96.2	2.8	3.0	1.2	Flotation Aid[d]
					6.0			After 12 hours holding
Hatboro, Pa.	R.S.	7,300	300	96.0	4.0	2.95	0.8	Flotation Aid
Morristown, N.J.	R.S.	6,800	200	97.0	3.5	1.70	0.5	Standard
Omaha, Nebr.	R.S.	19,660	118	99.8	5.9	7.66	0.8	Flotation Aid
					8.8			After 24 hours holding
Omaha, Nebr.	M.L.	7,910	50	99.4	6.8	3.1	0.8	Flotation Aid
Belleville, Ill.	R.S.	18,372	233	98.7	5.7	3.83	0.4	Flotation Aid
Indianapolis, Ind.	R.S.	2,960	144	95.0	5.0	2.1	1.47	Flotation Aid
					7.8			After 12 hours holding
Warren, Mich.	R.S. M.L.	6,000	350	95.0	6-9	5.2	1.75	Flotation Aid
Frankenmuth, Mich.	M.L.	9,000	80	99.1	6-8	6.5	1.3	Flotation Aid
Oakmont, Pa.	R.S.	6,250	80	98.7	8.0	3.0	1.0	Flotation Aid
Columbus, Ohio	R.S.	6,800	40	99.5	5.0	3.3	1.0	Flotation Aid
Levittown, Pa.	R.S.	5,700	31	99.4	5.5	2.9	1.0	Flotation Aid
Nassau Co., N.Y. Bay Park S.T.P.	R.S.	8,100	36	99.6	4.4	4.9	1.2	Flotation Aid
Nassau Co., N.Y. Bay Park S.T.P.	R.S.	7,600	460	94.0	3.3	1.3	0.33	Standard
Nashville, Tenn.	R.S.	15,400	44	99.6	12.4	5.1	0.66	Flotation Aid

[a] M.L. - Mixed liquor from aeration tanks.

[b] R.S. - Return sludge.

[c] Standard - Indicates no flotation aid and no holding before sampling.

[d] Flotation Aid - Indicates use of coagulant-flotation aid.

- A better solids capture (~ 97 percent)
- Maintenance of sludge in an aerobic condition
- Higher operating cost (power and flotation aids)

Since DAF effluent samples often have an 8 to 10 mg/l dissolved oxygen (DO) concentration, odor and phosphorus release problems seldom develop. Samples analyzed for total phosphate have given average subnatant values of 14 mg/l with average feed values of 570 mg/l. Total phosphate in the float in these samples averaged 1,400 mg/l.

4.3.4 Components of a Typical Flotation Unit System

Figure 4-11 is a schematic flow diagram for a typical DAF thickener. The normal system consists of the following items:

- Sludge Feed Pump (or Flow Control) and Meter
- Polymer Mix and Feed Equipment
- Flotation Thickener
- Electrical Control Panel
- Air Compressor and Receiver
- Thickened Sludge Pump

4.3.5 Design and Performance of DAF Thickeners

It is recommended that wherever possible, laboratory and pilot scale tests be carried out on the sludge which is being considered for flotation thickening. Equipment and chemical suppliers routinely supply methods and assistance in this area. Wood and Dick [18] present an excellent description of test procedures. In the absence of sludge for test purposes, procedures recommended by Jones [19] should prove satisfactory for conventional sludges. It is necessary to know the total pounds of waste sludge, the design loading of the unit, and the operating cycle in hours per week to size flotation units.

Performance is dependent on solids loading, concentration, hydraulic loading, and removal efficiency, and a discussion of each follows:

- *Solids Loading*–A design solids loading of 2 lb/hr/ft^2 is frequently selected. This rate is achieved with the use of flotation aids and with or without the use of auxiliary recycle, depending upon the operating situation. Although the guaranteed

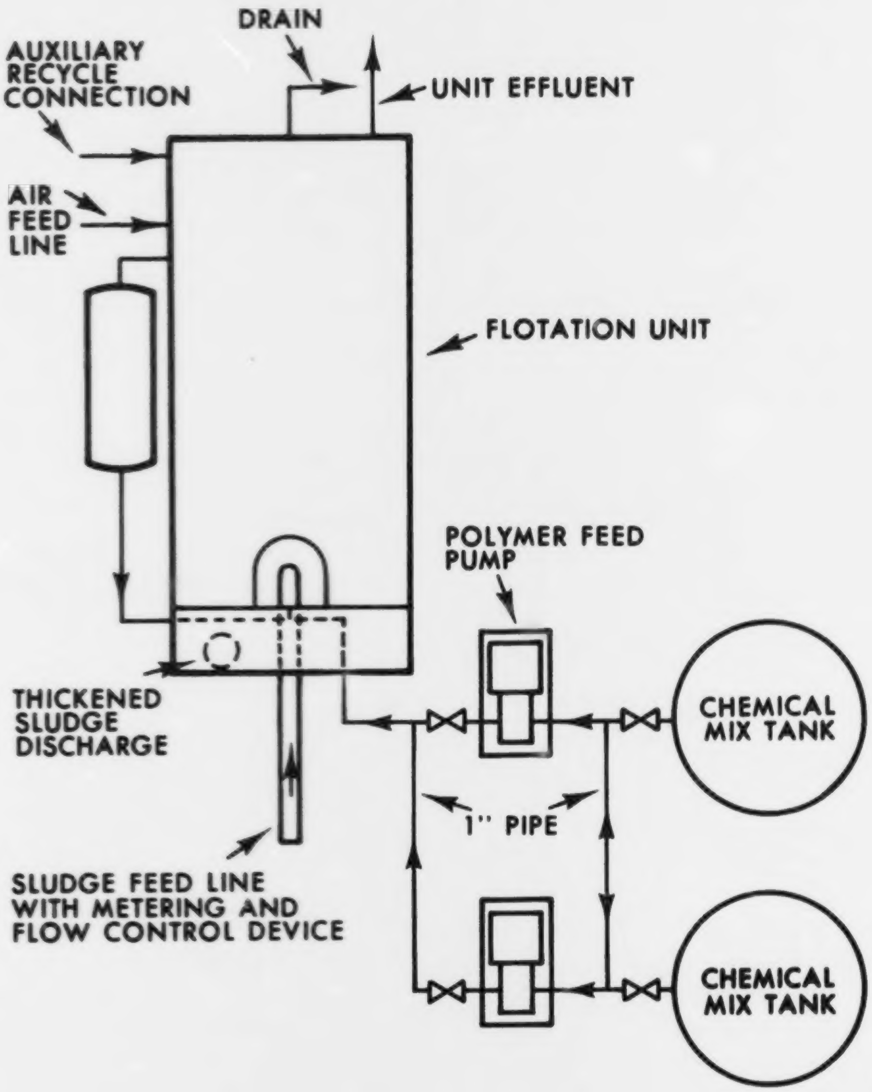

DRAIN
AUXILIARY RECYCLE CONNECTION
UNIT EFFLUENT
AIR FEED LINE
FLOTATION UNIT
POLYMER FEED PUMP
THICKENED SLUDGE DISCHARGE
CHEMICAL MIX TANK
1" PIPE
SLUDGE FEED LINE WITH METERING AND FLOW CONTROL DEVICE
CHEMICAL MIX TANK

FIGURE 4-11. Typical DAF thickener system.

loading with polymers is 2.0 lb/hr/ft², operating data indicate that in most cases 3.0 lb/hr/ft² can be expected. (The flotation thickener has a built-in capacity for 4.0 to 5.0 lb/hr/ft², which allows for a large safety factor and flexibility in operation.) There are times when flotation can be accomplished without flotation aids, but generally auxiliary recycle is then employed. Loadings are generally less than 50 percent of those attainable with aids, and solids removals are questionable. Existing data show 50 to 80 percent solids recovery at various plants and times when flotation aids were not employed.

- *Concentration*–A 4 percent minimum float solids concentration by weight is normally specified for design purposes. However, a 5 to 6 percent float solids concentration can be expected. Flotation without chemical aids generally results in a solids concentration that is about 1 percentage point less than with flotation aids. Further concentration is possible in a holding tank. While float density depends upon the nature of the sludge and the amount of air entrained, tests have shown that a weight of about 6 lb per gal can be considered for sizing of handling and storage facilities. After a sludge is held for several hours, the air is dissipated, and the sludge possesses a more typical density, in addition to being further concentrated.

- *Hydraulic Loading*–The maximum hydraulic loading or overflow rate is set at 0.80 gpm/ft². The minimum solids concentration compatible with this hydraulic loading and a solids loading of 2 lb/hr/ft² is 5,000 mg/l. Lesser solids levels or higher hydraulic loadings result in lower efficiencies and/or float solids concentrations.

- *Removal Efficiency*–Tests have shown that at least a 95 percent removal of suspended solids can be expected with the flotation unit when flotation aids are employed.

The appropriate size of a flotation unit for an existing plant can be calculated from the solids loading, the sludge settling characteristics, and the solids concentration of the sludge. If the thickening system is to be sized for a new plant, the following should be considered:

- The population equivalent.

- A design loading of 0.20 lb/capita/day of dry solids to the treatment plant.

- The quantity of wastewater solids which reach the activated sludge process; that is, the percent of the solids that are not removed in primary treatment. Based on a conventional activated sludge plant with primary settling, a design loading of 0.20 lb/capita/day suspended solids, and 30 percent BOD removal in the primary, approximately 0.085 lb/capita/day of excess activated sludge solids will be produced. To be conservative, 0.10 lb/capita/day EAS production can be assumed.

For a modified activated sludge system (with no primary settling) 0.17 lb/capita/day EAS solids production may be used unless a high BOD is expected. If unusually high BOD loadings are anticipated, additional calculations should be performed to estimate the quantity of sludge. For aerobically digested sludges, it may be assumed that 50 percent of the volatile solids are destroyed and this causes the previously estimated sludge quantities to be reduced by 35 percent. Where operating records are available in existing plants, the amount of excess sludge should be calculated from this information.

- Use a solids loading of 2.0 lb/hr/ft^2 for the flotation unit.

- Assume the number of hours per week which the unit will be operated. Although a flotation unit does not require continuous operator attention, periodic attention to the system should be scheduled. Generally, for plants less than 2 mgd, a 40 hr/wk schedule is adequate. For plants 2 to 5 mgd, 80 hr/wk (2 shifts, 5 days) may be realistic. For plants ranging from 5 to 20 mgd, 100 hr/wk (Monday a.m.-Friday p.m.) may be used. In plants larger than 20 mgd, it is expected that operators will be on duty 24 hr/day, 7 days per week. Of course, any information from an expected operating schedule must be considered.

- Check the maximum hydraulic loading of 0.80 gpm/ft^2.

Design Example

Population equivalent is 140,000 persons. Estimated plant size would then be 14 mgd. Since the plant is conventional activated sludge and has primary settling tanks that remove 50 percent of the influent solids, only 50 percent of the solids will reach the activated sludge process. The EAS to be handled is approximately 0.10 lb/capita/day or:

$$140{,}000 \times 0.10 \times 7 = 98{,}000 \text{ lb/wk}$$

For a loading of 2 lb/hr/ft^2 and a 100 hr/wk operating schedule the size unit required is:

$$\frac{98{,}000 \text{ lb/wk}}{2 \text{ lb/hr/ft}^2 \times 100 \text{ hr/wk}} = 490 \text{ ft}^2$$

Use either two 250 ft^2 units or one 500 ft^2 unit. It is generally preferable to apply two smaller prefabricated units rather than one large unit for obvious reasons of plant flexibility.

4.3.6 Costs

Stanley Consultants [16] have determined the DAF process cost given in Table 4-6.

TABLE 4-6

DAF THICKENING COSTS FOR VARIOUS PLANT SIZES

Plant Size (mgd)	Cost ($/ton Dry Solids)*		
	O & M	Amortization	Total
1	9.00	17.00	26.00
10	1.20	2.80	4.00
100	0.50	1.50	2.00

*Based on 1972 cost figures.

The costs are based on:

- Amortization at 7 percent for 20 years
- Labor rate of $6.25/hr
- Power cost of $0.01/kwh
- No chemicals
- Influent sludge with 0.5 percent solids concentration and a solids float of 3.5 percent
- Surface loading rate of 14.4 lb/day/ft^2
- Source: Envirotech and Stanley Consultants

Use of flotation aids would add another $2 to $7 per ton. In reviewing the above costs, due credit should be taken for the substantial benefits afforded by a consistent, effective, and highly operable method of sludge thickening, and consequent effects on the overall sludge treatment system operability and costs.

4.3.7 Integration of DAF Thickening into the Conventional Activated Sludge Plant

For a completely successful flotation system, as presented by Jones [19], close attention must be given to incorporation of the system into the activated sludge plant. Figure 4-12 indicates a suggested schematic flow diagram for incorporating a flotation thickener into the conventional activated sludge plant. It can be seen that the flotation thickener does not take the place of the final clarifier, but handles the excess activated sludge only. Flotation unit's feed is taken from the final settling tanks, while the flotation unit's subnatant is returned to the aeration tanks. Thickened sludge or float is delivered to a holding tank. Sludge holdup in the tank often provides further concentration through extended detention. Resulting supernatant can be returned to the flotation unit. The holding tank is not an additional expense to the DAF process, since it is necessary with any thickening device where sludge is to be delivered to a vacuum filter or a digester. The hoiding tank is equipped for mixing the gravity thickened primary and DAF thickened excess activated sludges. Such mixing appreciably improves dewatering of fresh solids on a vacuum filter. Air at a rate of 0.5 cfm per 100 gallons of storage is an effective mixer and prevents septicity. Decanting lines at various levels are advisable. Tank bottoms should be sloped at least 1.5:1.0. Tank volume generally affords 24 to 48 hours holding capacity.

Although flotation effluent is recycled during normal operation, auxiliary recirculation is provided from the primary tank effluent. Recirculation of primary tank effluent rather than the plant effluent minimizes the hydraulic loading and recirculated waters within the plant. However, if primary effluent containing more than 200 mg/l suspended solids or unusual amounts of stringy material is anticipated, plant effluent should be employed. It is important that feed to the flotation unit is fully controlled. In some cases where flight scrapers are used in the final settling tanks, wide variations in feed concentrations could result. The feed takeoff line to the flotation unit should be located to minimize fluctuations. Sludge wasting from a sludge reaeration tank is desirable. A feed pump with an off-on characteristic is definitely undesirable. A flow indicator and flow control devise must be located at the flotation unit control station. Depending upon the physical layout of the wastewater treatment plant, certain feed pumps and effluent pumps may be necessary for the flotation unit. It is generally considered acceptable that one pump be furnished to supply a controlled feed rate to the flotation unit. Unit effluent can be returned to the aeration tanks or plant influent wet well by gravity. Float from the unit can generally be delivered by gravity, to the holding tank, from which it is pumped either to a filter or a digester.

4.3.8 Effect of Oxygen Activated Sludge

There is some evidence that activated sludges from pure oxygen systems are more amenable to flotation thickening than activated sludges from conventional air systems. Pilot tests at Louisville, Kentucky, indicate that with a small polymer dose of about 3 lb/ton of dry solids, an influent solids concentration of 1.7 to 2 percent was increased to

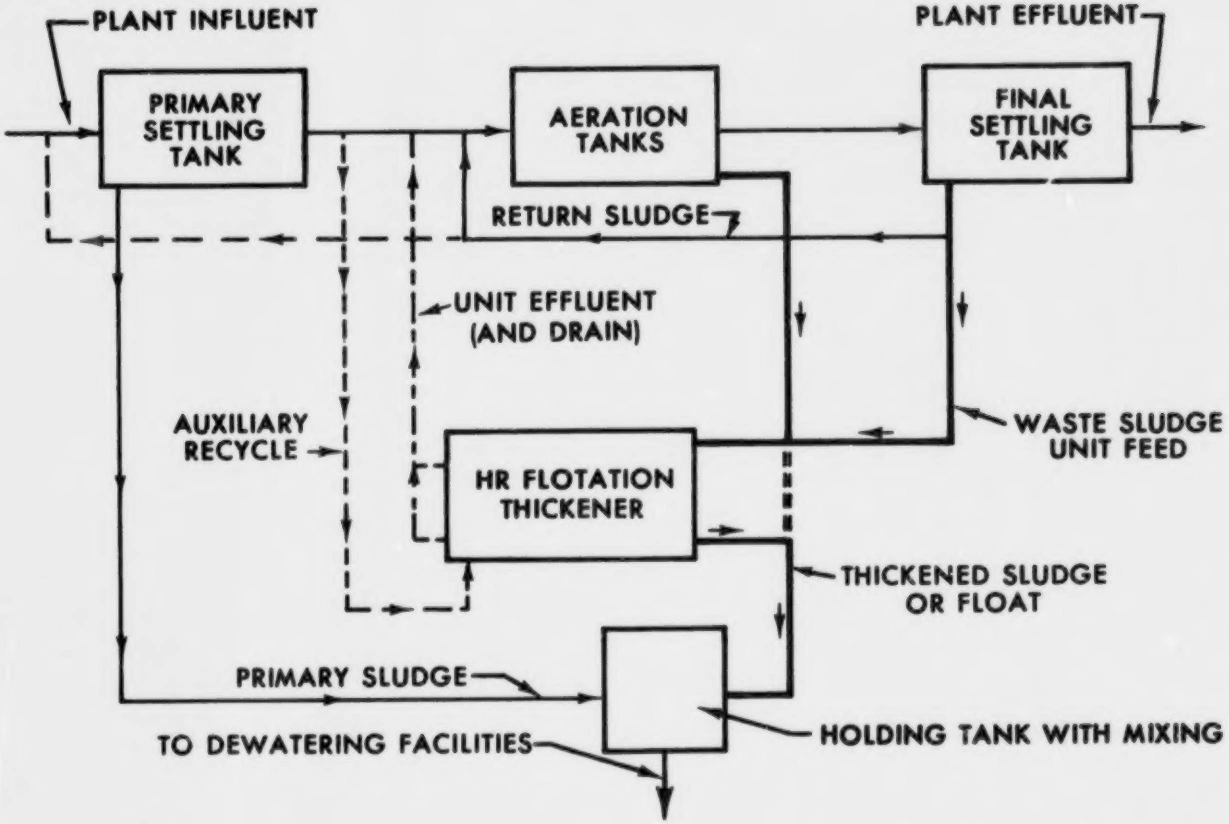

PLANT INFLUENT
PLANT EFFLUENT
PRIMARY SETTLING TANK
AERATION TANKS
FINAL SETTLING TANK
RETURN SLUDGE
UNIT EFFLUENT (AND DRAIN)
AUXILIARY RECYCLE
HR FLOTATION THICKENER
WASTE SLUDGE UNIT FEED
THICKENED SLUDGE OR FLOAT
PRIMARY SLUDGE
HOLDING TANK WITH MIXING
TO DEWATERING FACILITIES

FIGURE 4-12. Schematic flow diagram of a conventional activated sludge plant incorporating a DAF thickener.

between 6 and 7 percent solids at loading rates of 6 to 10 lb/hr/ft^2 and 200 percent recycle. The subnatant SS concentration varied from 80 to 570 mg/l and this represented a 97 to 99.7 percent solids recovery. Similar results have been observed on a plant scale at the Westgate plant in Fairfax County, Virginia.

The improved thickening and dewatering characteristics of oxygen activated sludge have been previously noted, and so the above findings are not surprising. However, as pointed out in Chapter 10, the Westgate plant has opted for gravity thickening of combined sludges instead of DAF thickening because their oxygen activated sludge responds to this type treatment. This illustrates the point that where the sludges respond properly to gravity thickening, that method is preferred. Unfortunately, few air system activated sludges respond well to gravity thickening.

4.4 Centrifugal Thickening

There has been limited use of centrifuges for thickening of EAS. The centrifugal thickening process can have substantial maintenance and power costs. It has been used only where space limitations or sludge characteristics make other methods unsuitable [20]. Further, if a particular sludge can be effectively thickened by gravity or by flotation thickening without chemicals, centrifuge thickening is not economically feasible [16]. The use of centrifuges for dewatering is discussed in Chapter 6. The solid bowl conveyor disc-nozzle and basket centrifuges have been evaluated for sludge thickening.

4.4.1 Solid Bowl Conveyor Type Centrifuge–Sludge Thickening

Ettelt and Kennedy [21] evaluated flotation along with disc and solid bowl centrifugal equipment on excess activated sludge at the Chicago Sanitary District's southwest plant and opted for flotation thickening. This was in spite of the fact that the solid bowl centrifuge, when processing activated sludge alone could thicken the sludge to from 6.6 to 7.5 percent. The principal difficulty with attempts to use the solid bowl conveyor type unit is reflected in work by Ooten and Miele [22]. In order to achieve a 90 percent solids capture and a 4 percent solids thickened activated sludge, the sludge had to be adequately conditioned with chemicals. The chemicals added an additional processing cost of approximately $20/ton. The solid bowl unit imposes the drastic effect of high shear on the fragile activated sludge and chemicals are required to prevent floc breakup. Further, the conveyability requirements of this type unit result in a high chemical demand.

4.4.2 Disc-Nozzle Centrifuge

The disc-type unit has concentrated activated sludge to about 7 percent when operated at 6,000 rpm. Operational problems, however, made its use impractical. These problems were caused by clogging of the sludge discharge nozzles, which required repeated maintenance. Rotary screens were found effective in removing a large amount of the oversized solids, but this resulted in low throughput. Despite these difficulties some field testing of modified equipment has prompted the selection of disc-nozzle centrifuges by three treatment plants. They are the Village Creek plant at Ft. Worth, Texas; the plant at Kelowna, British Columbia and Idaho Falls, Idaho. Construction has been completed and operation commenced at the Ft. Worth plant. Data from operation of their disc-nozzle centrifuge indicate that:

- Activated sludge is being concentrated from 4 to 5-1/2 percent solids.
- Both rotary screens and cyclones were installed to help alleviate plugging problems.
- Rotary type screens have recently been installed and their use, together with hot-water flushing will hopefully result in 2 to 3 weeks operation between cleanings.

A disc centrifuge has been successfully field tested for thickening waste activated sludge at an eastern Pennsylvania community. Using a 30-inch centrifuge with a 150-hp motor and 300 gpm feed rate, the disc centrifuge produced a 5 percent underflow with 90 percent solids recovery. Since the plant did not have primary treatment, it was necessary to install a screening device ahead of the centrifuge. The screening effectiveness was demonstrated in that the nozzles of the centrifuge did not plug. The combination of effective screening and patented recirculating system (allowing a larger nozzle size) was instrumental to the good performance. Normally, however, disc centrifuges are not recommended where activated sludge treatment has not been preceded by primary treatment.

Field test results have also been reported by Vaughn and Reitwiesner [23]. The system studied is illustrated in Figure 4-13. While it has been suggested that cyclones and screens only be used when primary clarifiers are not used recent field results indicate the continuous need for them. The effect of activated sludge SVI on solids capture at various thickened sludge solids levels is shown in Figure 4-14. As can be seen the thickened sludge solids content of 4 percent requires a SVI of about 100 to insure 90 percent solids capture. A small amount of flocculant may be required in some instances to assist both solids recovery and thickened sludge solids concentration.

An interesting comparison between gravity thickening of mixed sludge and a combination involving gravity thickening of primary sludge with centrifugal thickening of activated sludge is shown in Table 4-7.

The data again show the previously noted need for separate EAS thickening via an efficient process. However, the comparison would have been better if it had included flotation thickening.

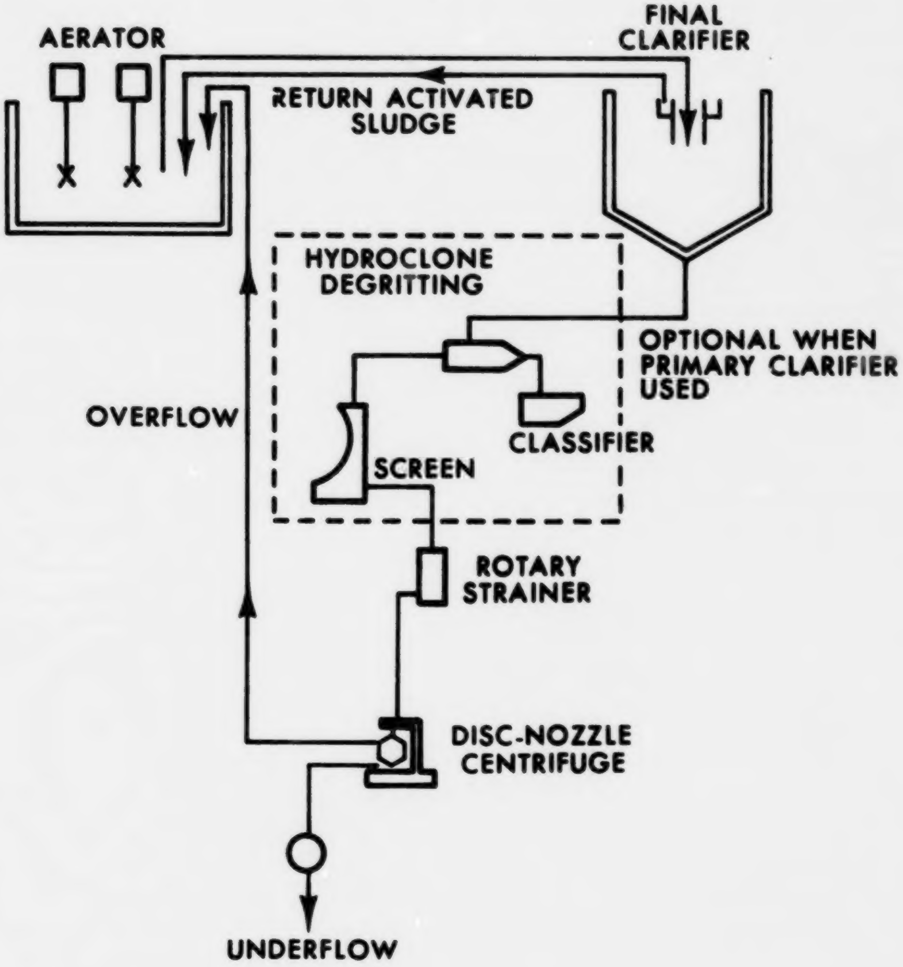

AERATOR
FINAL CLARIFIER
RETURN ACTIVATED SLUDGE
HYDROCLONE DEGRITTING
OPTIONAL WHEN PRIMARY CLARIFIER USED
OVERFLOW
CLASSIFIER
SCREEN
ROTARY STRAINER
DISC-NOZZLE CENTRIFUGE
UNDERFLOW

FIGURE 4-13. Thickening of activated sludge by disc-nozzle centrifuge [23].

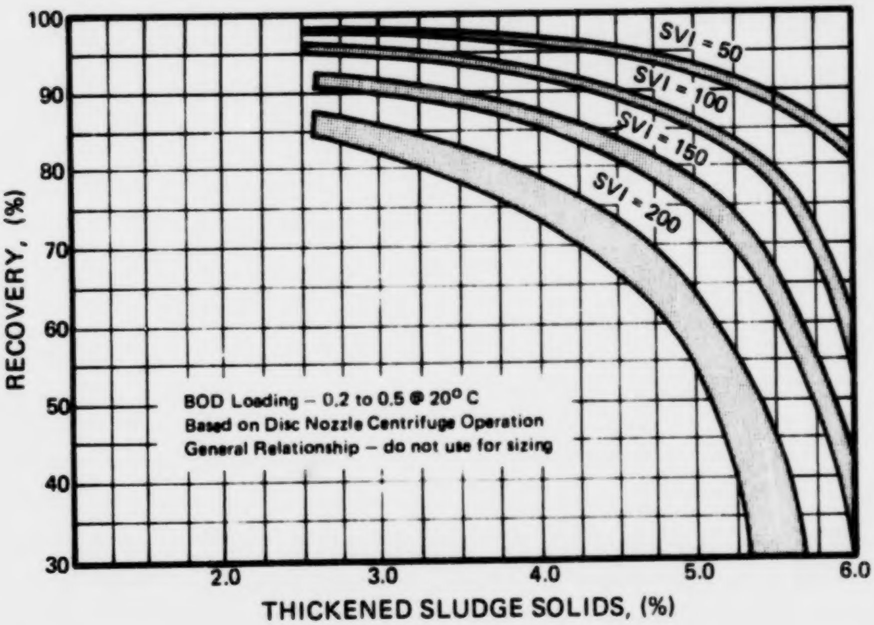


FIGURE 4-14. Effect of activated sludge settleability on capture and thickening [23].

TABLE 4-7

COMPARATIVE DATA ON TWO ALTERNATIVE SLUDGE THICKENING PROCESSES [23]

	Alternative No. 1	Alternative No. 2		
Parameter	Gravity Primary and Secondary	Gravity Primary	Disc-Nozzle Centrifuge Secondary	Weighted Total
Unit Area, ft^2/ton/day	250	104	6	65
Capital Cost, $/ton/day	6,300	3,150	5,900	4,260
Power cost ($/ton)	0	0	1.70	0.67
Solids concentration (%)	5	12	5	9.3
Solids recovery (%)	95	95	90	90+
Flexibility	None	–	–	Excellent

Note: $ft^2 \times 0.0929 = m^2$; ton (short) X 0.9078 = ton (metric)

4.4.3 Basket (Imperforate Bowl-Knife Discharge) Centrifuge

This unit has been extensively field tested at small plants for the thickening of excess activated sludge to a 9 to 10 percent solids level. This level of solids facilitates land disposal. The basket centrifuge's application is intermediate between normal thickening and what can be construed as dewatering. Since there are few actual plant installations, there is little available data.

4.4.4 Performance Data

Typical performance data for the disc, basket, and solid bowl centrifuges when they are employed in the thickening of EAS, are presented in Table 4-8. Note that chemical addition is not always required.

4.5 Sludge Blending

With the increased use of separate thickening processes for primary and activated sludges, more consideration of this subject is in order. Thickening makes both primary and excess

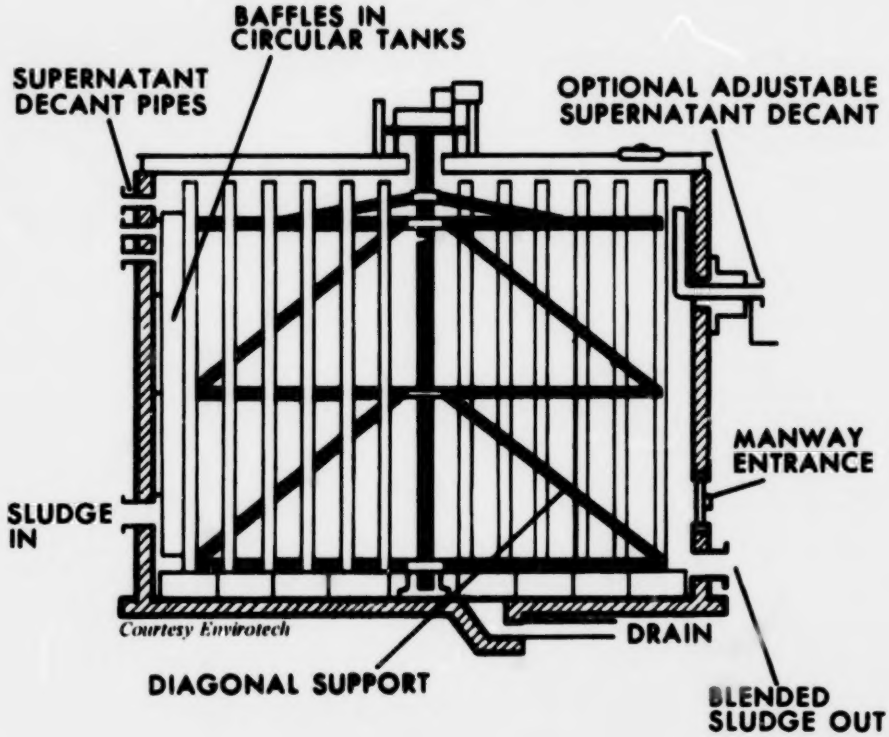
TABLE 4-8

CENTRIFUGAL THICKENING PERFORMANCE DATA [20]

Type of Sludge	Centrifuge Type	Capacity (gpm)	Feed Solids (%)	Underflow Solids (%)	Solids Recovery (%)	Polymer Requirement (lb/ton)
EAS	Disc	150	0.75-1.0	5-5.5	90+	None
EAS	Disc	400	--	4.0	80	None
EAS (after Roughing Filter)	Disc	50- 80	0.7	5-7	93-87	None
EAS (after Roughing Filter)	Disc	60-270	0.7	6.1	97-80	None
EAS	Basket	33-70	0.7	9-10	90-70	None
EAS	Solid-Bowl	10-12	1.5	9-13	90	--
EAS	Solid-Bowl	75-100	0.44-0.78	5-7	90-80	None
EAS	Solid-Bowl	110-160	0.5 -0.7	5-8	65	None
					85	<5
					90	5-10
					95	10-15

activated sludge more difficult to blend. This results in the need for mechanical mixing in sludge holding tanks. This problem has been recognized by some suppliers and engineering firms and is well known to plant personnel. An example of the type unit which can alleviate problems in blending thickened sludges is shown in Figure 4-15.

Contact with suppliers of mixing equipment should be made when considering sludge blending facilities. It is also probable that in-line mixers are of interest and should be considered. These units are designed to promote blending at minimum shear. Bubbling air through various sludge mixtures (a previous practice) for the purpose of blending has been found to be ineffective with thickened sludges. Use of some air injection as well as minimum holding time should be considered to prevent septicity.


BAFFLES IN CIRCULAR TANKS
SUPERNATANT DECANT PIPES
OPTIONAL ADJUSTABLE SUPERNATANT DECANT
MANWAY ENTRANCE
SLUDGE IN
DRAIN
DIAGONAL SUPPORT
BLENDED SLUDGE OUT

Courtesy Envirotech

FIGURE 4-15. Eimco sludge storage tank blender mixer.

4.6 References

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CHAPTER 5

SLUDGE STABILIZATION (REDUCTION)

5.1 Functions, Methods, and Occurrences

The principal purposes of stabilization are to make the treated sludge less odorous and putrescible, and to reduce the pathogenic organism content. Some procedures used to accomplish this objective can also result in other basic changes in the sludge. The selection of a certain method hinges primarily on the final disposal procedure planned for the sludge. If the sludge is to be dewatered and incinerated, frequently no stabilization procedure is employed. Most stabilization methods, particularly anaerobic and aerobic digestion, result in a substantial decrease in the amount of suspended sludge solids. Hence, the corollary function of reduction is included in the description of these processes.

Both anaerobic and aerobic digestion are currently increasing in popularity. The former is receiving revived attention because of the potential benefits of methane production, the energy shortage, increasing realization that many of the previous problems experienced were due to other wastewater process considerations, and the emphasis on final disposal on land. Interest in aerobic digestion of excess activated sludge is growing because it has the potential for providing a good quality liquid process stream and can produce exothermic reaction conditions. Composting is being practiced in several United States cities and is being actively investigated for others. A major impetus for processes such as anaerobic and aerobic digestion, lime treatment, and composting is the growing emphasis on utilization of sludge rather than mere disposal. Chlorine oxidation is of limited use for special situations or where septic tank wastes are involved. Heat treatment has been installed in several new United States plants to improve sludge conditioning and dewatering economics.

5.2 Anaerobic Digestion

Excellent descriptions of this oxygen devoid process appear in the literature [1,2]. Table 5-1 illustrates the biochemical reactions occurring in anaerobic digestion.

Digestion is a complex biochemical process in which several groups of anaerobic and facultative organisms simultaneously assimilate and break down organic matter. For purposes of simplification, it is a two-phase process and can be described as follows:

1. In the first phase facultative, acid-forming organisms convert the complex organic substrate to volatile organic acids. Acetic, propionic, butyric, and other organic acids are formed. In this phase little change occurs in the total amount of organic material in the system, although some lowering of pH results. Alkaline buffering materials are also produced.

TABLE 5-1

ANAEROBIC DIGESTION–BIOCHEMISTRY [1]

Raw Sludge	+	Micro-organisms "A"	K_1 $\rightarrow$	Nonreactive Products	+	Reactive Products	+	Micro-organisms "B"	K_2 $\rightarrow CH_4 + CO_2$	+	Other end Products
Complex substrate		Principally acid formers		CO_2, H_2O		Organic acids		Methane fermenters			H_2O, H_2S
Carbohydrates, Fats, and Proteins				Stable and intermediate degradation products		Cellular and other intermediate degradation products					Cells and stable degradation products
				Cells							

2. The second phase involves conversion of the volatile organic acids to primarily methane and carbon dioxide.

This anaerobic process is essentially controlled by the methane-producing bacteria. These bacteria grow at a relatively low rate and have generation times which range from slightly less than 2 days to about 22 days. Methane formers are very sensitive to pH, substrate composition, and temperature. If the pH drops below 6.0, methane formation ceases, and there is no decrease in organic content of the sludge. The methane bacteria are highly active in the mesophilic and thermophilic ranges. The mesophilic range is between 80° F and 110° F, while the thermophilic range is between 113° F and 149° F. Essentially all digesters in the United States operate within the mesophilic temperature range. Garber [3] has recently experimented with thermophilic temperatures at the city of Los Angeles Hyperion Plant. The main advantage was found to be that of improved dewatering. The cost of heating to this temperature was justified by the increased efficiency.

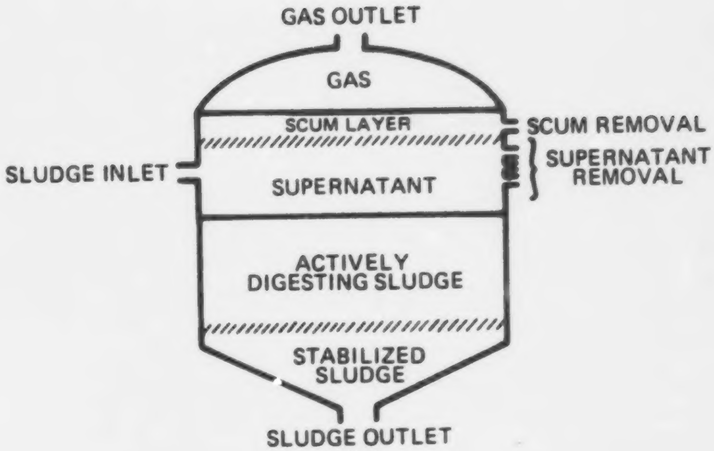
5.2.1 Types of Anaerobic Digestion Systems

The standard rate and high rate systems are the two main digestion processes employed. Schematics of the processes as well as their operating criteria are given in Figure 5-1.

In practice, four types of systems have evolved from the two basic digestion modes. They are:

1. Standard Rate Digestion–One Stage

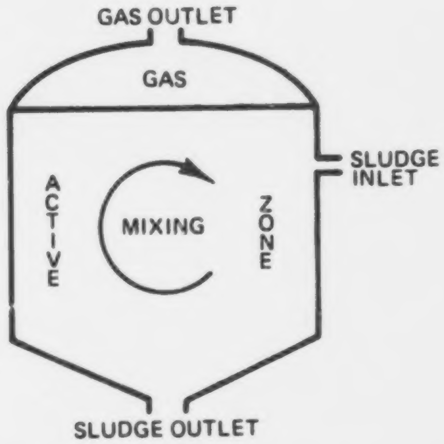
2. High Rate Digestion–One Stage


GAS OUTLET
GAS
SCUM LAYER
SCUM REMOVAL
SUPERNATANT REMOVAL
SLUDGE INLET
SUPERNATANT
ACTIVELY DIGESTING SLUDGE
STABILIZED SLUDGE
SLUDGE OUTLET

(A)

STANDARD RATE DIGESTION

1. UNHEATED
2. DETENTION TIME 30 – 60 DAYS
3. LOADING 0.03 – 0.10 lb. VSS/cu. ft./day
4. INTERMITTENT FEEDING AND WITHDRAWAL
5. STRATIFICATION


GAS OUTLET
GAS
SLUDGE INLET
ACTIVE
MIXING
ZONE
SLUDGE OUTLET

(B)

HIGH RATE DIGESTION

1. HEATED TO 85° – 95° F
2. DETENTION TIME 15 DAYS OR LESS
3. LOADING 0.10 – 0.50 lb. VSS/cu. ft./day
4. CONTINUOUS OR INTERMITTENT FEEDING AND WITHDRAWAL
5. HOMOGENEITY

FIGURE 5-1. Standard rate and high rate digestion.

3. Two-Stage Digestion

4. Anaerobic Contact Process

In the standard rate, one-stage digestion process as shown in Figure 5-1(A), fresh sludge is usually added to the system two or three times daily. As decomposition proceeds, three distinct zones develop. A scum layer is formed at the top of the digester, and beneath it are supernatant and sludge zones. The sludge zone has an actively decomposing upper layer and a relatively stabilized bottom layer. The stabilized sludge accumulates at the base of the digester. Supernatant is usually returned to the influent of the treatment plant and this practice can create problems and reduce overall treatment plant efficiency even in smaller plants.

The high rate, one-stage system (Figure 5-1(B)) requires a separate postdigestion thickening process if dewatering is practiced. This type system is increasingly being used in plants featuring anaerobic digestion because of the beneficial aspects of mixing, improved process control, and lack of in-tank settling problems. Two-stage digestion is shown in Figure 5-2.

The two-stage process can operate at various loading rates and therefore is not always clearly defined as being either standard or high rate. It evolved as an attempt to provide additional gas production as well as a separate settling and thickening process in the secondary digester. The process is successful when primary sludge or combinations of primary sludge and limited amounts of secondary sludges constitute the system's feed. With the advent of wastewater treatment systems that are more efficient than simple sedimentation, large quantities of activated and sometimes advanced waste treatment (AWT) sludges are produced at the plants. This additional sludge, when placed in a two-stage anaerobic digestion process, can cause high operating costs and poor plant efficiencies. The basic cause of the problem is that the additional solids do not readily settle after digestion. Typical resultant sludge processing problems are described in Chapter 10.

Schroepfer and Ziemke [4] and McCarty [5] have fully discussed the anaerobic contact process and it is shown in Figure 5-3. In the process operation, sludge from a high rate digester is settled in a second-stage digester. The second-stage digester operates as a settling basin to permit removal of microorganisms from the effluent. The organisms, as in the activated sludge process, are returned to the digester and seed the raw waste. This process has an increased rate of waste decomposition, when it is compared with high rate digestion.

5.2.2 Design Criteria

Anaerobic digestion is influenced by both physical and chemical factors, some of which are listed in Table 5-2.

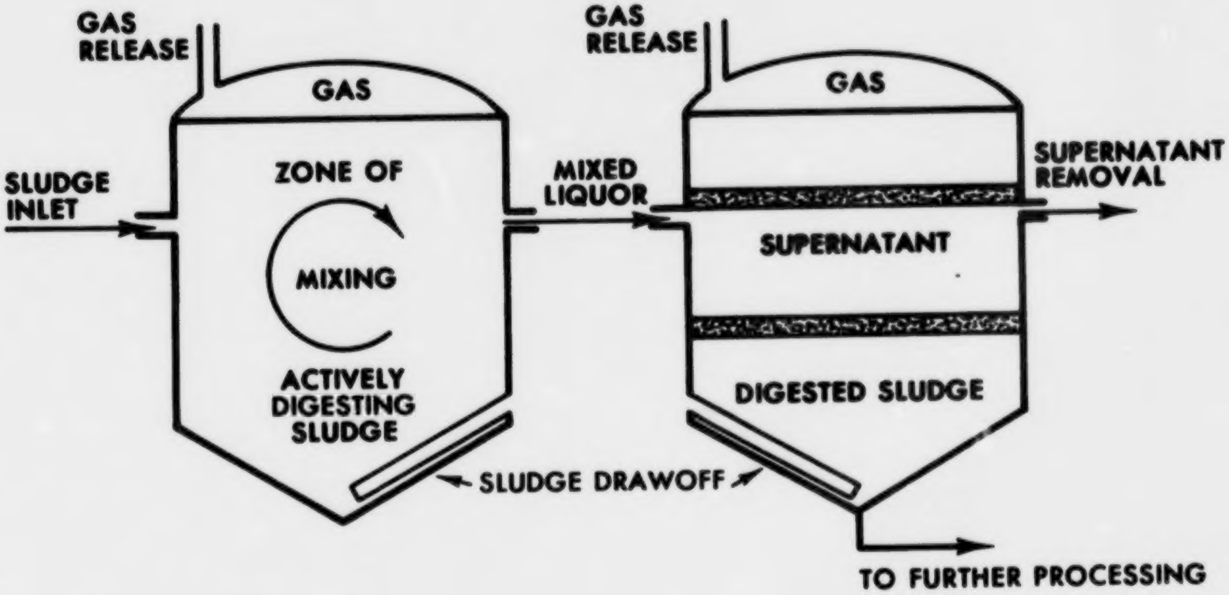

GAS RELEASE
GAS
SLUDGE INLET
ZONE OF
MIXING
ACTIVELY DIGESTING SLUDGE
MIXED LIQUOR
SLUDGE DRAWOFF
GAS RELEASE
GAS
SUPERNATANT REMOVAL
SUPERNATANT
DIGESTED SLUDGE
TO FURTHER PROCESSING

FIGURE 5-2. Two-stage anaerobic digestion.

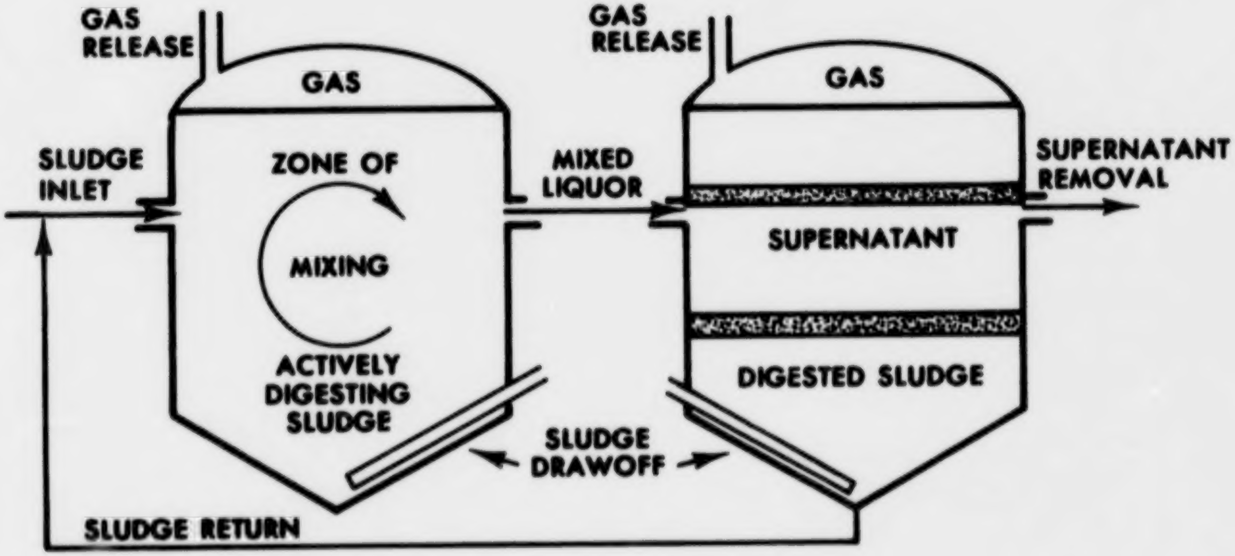

GAS RELEASE
GAS
SLUDGE INLET
ZONE OF
MIXING
ACTIVELY DIGESTING SLUDGE
MIXED LIQUOR
GAS RELEASE
GAS
SUPERNATANT REMOVAL
SUPERNATANT
DIGESTED SLUDGE
SLUDGE DRAWOFF
SLUDGE RETURN

FIGURE 5-3. Anaerobic contact digestion [2].

TABLE 5-2

PHYSICAL AND CHEMICAL FACTORS

Physical Factors	Chemical Factors
Detention Times	pH
Temperature	Alakalinity
Solids Concentration	Volatile Acid Content
Degree of Mixing	Nutrients
Solids Loading and Distribution	Toxic Materials

The most important operational factors controlling the design of an anaerobic digestion system are the combined effects of temperature mixing and the biological solids retention time (SRT). The rate of bacterial growth and, therefore, the rate of stabilization increases and decreases with temperature within certain limits [6]. Figure 5-4 shows the effects of temperature on digestion time.

Time values were omitted since the digestion period is also affected by several other factors such as pH, bacterial population, mixing, rate of feeding, and sludge characteristics. Field investigations of high rate digestion in a controlled temperature environment have indicated that high efficiencies are possible at liquid retention times as low as 2 days, so long as the SRT is equal to or greater than some critical time. This critical solids retention time (SRTc) is the time period below which digestion falls as a result of washout of the slow-growing methane formers. The regeneration rate for the slowest methane formers is about 10 days (SRTmin) at 95° F [6]. When the digestion time is decreased below SRTmin, decomposition of volatile organics is slowed until complete failure occurs at an SRTc of about 3 to 4 days.

Factors influencing the SRT are the volatile solids loading on the digester, the volatile percentage in the total suspended solids, and the suspended solids concentration in the raw sludge. A series of curves showing the relationship among solids loading, solids retention time, and sludge solids is presented in Figure 5-5. The volatile solids loading to the digester should always be adjusted, based on the volatile solids concentration in the sludge, so that a detention time above the SRTmin is maintained. For a volatile solids loading of 0.1 lb/cu ft/day (see Figure 5-5), SRTc of 10 days would be met by all sludges containing volatile solids concentrations greater than 1.5 percent.

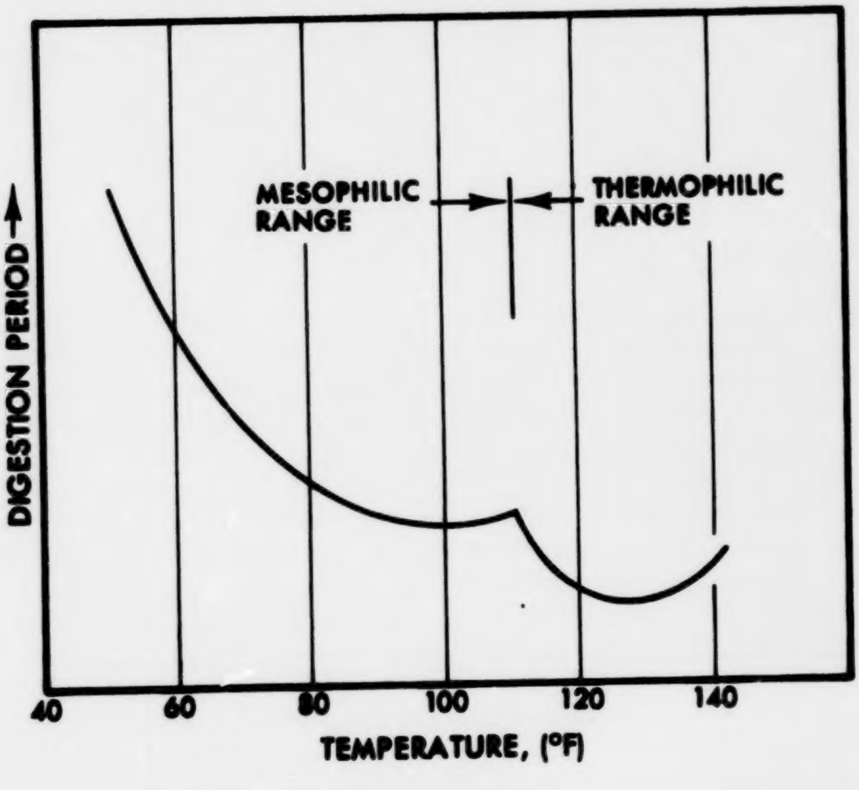


FIGURE 5-4. Influence of temperature on digestion time [7].

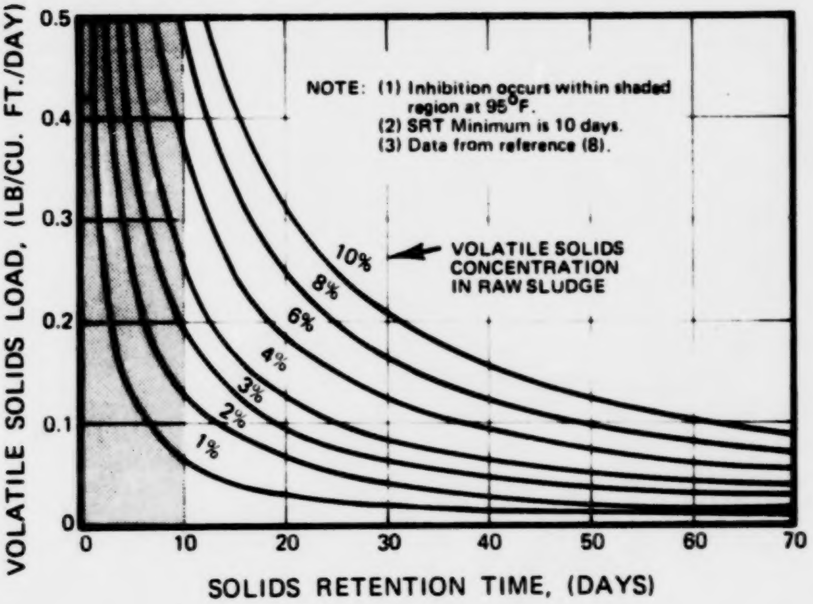


FIGURE 5-5. Plot of volatile solids loading vs. SRT for various feed solids.

Typical design criteria for anaerobic digesters are given in Table 5-3.

TABLE 5-3

TYPICAL DESIGN CRITERIA FOR STANDARD RATE AND HIGH RATE DIGESTERS [9]

Parameter	Low Rate	High Rate
Solids Retention Time (SRT), days	30 to 60	10 to 20
Solids Loading, lb VSS/cu ft/day	0.04 to 0.1	0.15 to 0.40
Volume Criteria, cu ft/capita		
Primary Sludge	2 to 3	1-1/3 to 2
Primary Sludge + Trickling Filter Sludge	4 to 5	2-2/3 to 3-1/3
Primary Sludge + Waste Activated Sludge	4 to 6	2-2/3 to 4
Combined Primary + Waste Biological		
Sludge Feed Concentration, percent solids (dry basis)	2 to 4	4 to 6
Digester Underflow Concentration, percent solids (dry basis)	4 to 6	4 to 6

As noted, the high rate process requires considerably less detention time, and volume, and operates successfully with a higher solids loading when compared to the conventional process. This is attributed to the greater use of the digestion tank for biological activity and improved mixing.

5.2.3 Process Control Considerations

Biochemical reaction conditions require close control for successful digestion and excellent descriptions of the problems and measures required to correct them can be found in the literature [10].

pH

Close pH control is necessary because methane bacteria are extremely sensitive to slight changes in pH. While the pH is usually allowed to vary from 6.6 to 7.4, it is generally wise to maintain the pH close to 7.0. In an anaerobic digester, a great quantity of carbon dioxide is produced during methane fermentation. The pH is, however, normally maintained by a bicarbonate buffer system. Figure 5-6 prepared by McCarty [11] illustrates the relationship among pH, the bicarbonate alkalinity of the digester liquor, and the fraction of CO_2 in the digester gas.

Because pH control is so important in digester operation, the dynamic nature of buffer destruction and formation in the digester should be understood. Therefore, this process is reviewed in the following equations for simple carbohydrates such as glucose.

$$C_6H_{12}O_6 \xrightarrow{\text{acid formers}} 3\,CH_3COOH$$

$$3\,CH_3COOH + 3\,NH_4HCO_3 \longrightarrow 3\,CH_3COONH_4 + 3\,H_2O + 3\,CO_2$$

$$3\,CH_3COONH_4 + 3\,H_2O \xrightarrow{\text{methane bacteria}} 3\,CH_4 + 3\,NH_4HCO_3$$

The first equation represents the breakdown of glucose to acetic acid by acid-forming bacteria. The acid is then neutralized, as shown in the second equation, by the bicarbonate buffer. If sufficient buffer is not present, the pH will drop, and the conversion of acetate to methane, as shown in the third equation, would be inhibited. The buffer consumed in the second reaction is reformed in the third reaction. In a properly operating digester a dynamic equilibrium is maintained between buffer formation and destruction, however, when an upset occurs, it is usually the methane bacteria which are adversely affected rather than the acid formers. Therefore, in an upset, net buffer consumption takes place, and the process is in danger of pH failure. When this happens, an external source of alkalinity must be supplied to maintain pH in the proper range.

Figure 5-6 indicates that the bicarbonate alkalinity should be maintained at a minimum level of 1,000 mg/l as $CaCO_3$ to ensure adequate pH control. To determine the bicarbonate alkalinity, both the volatile acid concentration and the total alkalinity must be measured. The bicarbonate alkalinity is then calculated as shown:

$$\text{Bicarbonate Alkalinity} = (\text{Total Alkalinity} - 0.8\ \text{Volatile Acids})$$

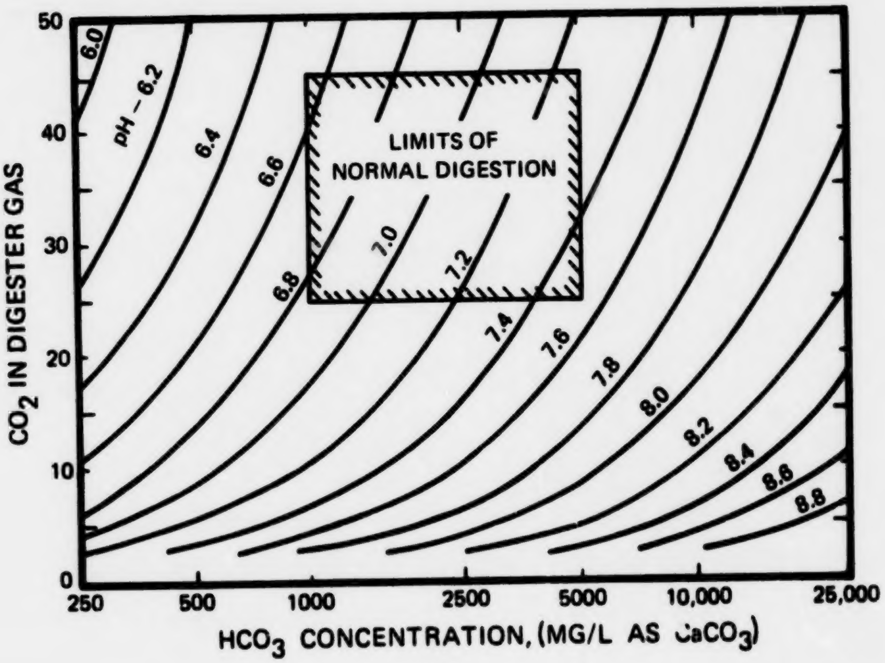


FIGURE 5-6. Relationship between pH and bicarbonate concentration [11].

The 0.8 factor in the above equation is required to convert the volatile acid units from mg/l as acetic acid to mg/l as $CaCO_3$, the equivalent alkalinity unit. The volatile acid to total alkalinity ratio should be maintained below 0.5 for good digester operation.

Temperature

The temperature response of methane bacteria is similar to that of other bacterial groups. Digestion of wastewater sludge is almost always conducted in the mesophilic range and the optimum temperature in this range is 95° F. More important than maintenance of a particular temperature is maintenance of the chosen temperature for operation at a constant value. A temperature change of 2 or 3 degrees can be sufficient to disturb the dynamic balance between the acid and methane formers. Such a disturbance will lead to an upset because the acid formers are able to respond more rapidly to changes in temperature than are the methane bacteria.

Nutrients

Little knowledge is available on the nutritional requirements of methane bacteria and this has been a stumbling block in the application of anaerobic treatment to industrial wastewaters. Speece and McCarty [12] have reported the most definitive work on the macronutrient and micronutrient requirements of these organisms. They indicate that domestic wastewater appears to contain all of the nutrients required. Thus, difficulty should only occur in digestion when a large fraction of the sludge is of industrial origin.

Potentially Toxic Materials

A review of this subject has been provided by Kugelman and Chin [13]. They indicate that toxicity, in general, can be due to an excessive quantity of any material, even a substance normally considered a nutrient. The concentration at which a substance starts to exert a toxic effect is difficult to define because it can be modified by antagonism, synergism, and acclimation. In addition, the organic loading and biological solids retention time can cause a stress on the process and this stress can affect toxicity. The substances which can produce toxicity when present in municipal sludge in an excessive concentration, include heavy metals, sulfides, surface active agents, light metals, and certain organics. All of these can gain entrance to wastewater sludge from industrial sources. In addition, light metal cations will enter sludge if an alkaline material is added to control the pH. Several papers [13,14,15] review the best engineering data available on toxicity. While reference should be made to these papers for complete information, general information on some substances is given in Table 5-4.

TABLE 5-4

SUBSTANCES AND CONCENTRATIONS CAUSING TOXICITY IN WASTEWATER SLUDGE DIGESTION [10]

Substance	Concentration (mg/l)
Sulfides	200
Soluble Heavy Metals	>1
Sodium	5,000 - 8,000
Potassium	4,000 - 10,000
Calcium	2,000 - 6,000
Magnesium	1,200 - 3,500
Ammonium	1,700 - 4,000
Free Ammonia	150

It must be emphasized that the values in this table are only guides. If toxicity is suspected, a thorough determination of the chemical constituents in the sludge should be made before definite conclusions are drawn. Potential solutions to toxicity problems, other than elimination of the chemical from the wastewater, should be evaluated in laboratory or pilot digesters.

Process Kinetics

Lawrence and McCarty [16] have reviewed the kinetics of anaerobic digestion and show that the overall process kinetics are controlled by methane bacteria. In addition, they found that the removal efficiency could be characterized by the equation:

$$E = \frac{S_o - \dfrac{K_g (1 + K_d - SRT)}{SRT \cdot K_m - (1 + K_d \cdot SRT)}}{S_o}$$

where

E = Substrate removal efficiency expressed in decimal form

S_o = Influent substrate concentration

K_g, K_d, and K_m = Kinetic constants

SRT = Biological solids retention time

The engineer and/or plant operator primarily controls the SRT. Thus, this becomes the fundamental design and control parameter. SRT is analogous to the sludge age parameter used in activated sludge system design. For a digestion system without sludge recycle, the SRT is numerically equal to the hydraulic retention time (HRT). This analysis points up a fallacy in current digestion criteria. Digesters are designed at present on the basis of either the volume per individual served, the weight of volatile solids per unit volume of digester per unit time, or HRT. Of these, the only valid criterion is HRT.

Values for the kinetic constants discussed above were determined experimentally by Lawrence and McCarty [17]. These values indicate that at 95° F the absolute minimum SRT for anaerobic digestion is three to four days. This value agrees well with the minimum HRT determined by Torpey [18] in field studies. For design purposes, a longer HRT should be utilized to provide a safety factor against upsets and to allow for fluctuations in sludge volume. In addition, the rate limiting step in some situations is solubilization of grease and/or protein, which requires HRT values longer than four days. Suggested retention times for high rate digesters were given in Table 5-3.

5.2.4 Process Performance Data

Anaerobic digestion treats sludge by converting approximately 50 percent of the organic solids to liquid and gaseous forms. Discussion of expected digester performance can best be illustrated by an example. The Chicago Sanitary District [19] digesters were designed as shown in Table 5-5. Note that the system design is that of high rate and must handle a high percentage of EAS. The summary of a year's operating results for the Southwest Treatment Plant's digesters is shown in Table 5-6. The data are for the period from July, 1964, to July, 1965, and were obtained at an average loading of 0.081 lb/day/cu ft of

TABLE 5-5

DESIGN DATA FOR CHICAGO DIGESTERS

Parameter	Value
Solids to digesters, tons/day	100
Sludge to digesters, % solids	3.3
Sludge to digesters, % volatile	67
Digester displacement, days	14
Volatile solids digested, %	40 to 45
Gas per pound volatile solids digested, cu ft	16 to 18
Heat per cubic foot of gas, BTU	600 to 650
Temperature of digesters, ° F	90 to 95
Sludge to digesters:	
Activated sludge solids, %	80 to 100
Primary sludge solids, %	20 to 0

TABLE 5-6

SUMMARY OF SOUTHWEST TREATMENT PLANT DIGESTER OPERATION

	Dry Solids %				Dry Solids (ton/day)			
	Total	Volatile	pH	Alkalinity (mg/l as $CaCO_3$)	Total	Volatile	Volatile Solids Reduction %	Gas Produced cu ft/lb Volatile Destroyed
Feed	3.1	65.3	6.5	671	82.9	54.4	--	--
Draw-off	2.4	54.7	7.1	2,162	63.4	35.0	35.7	20.0

volatile solids, a detention time of 16 days, and temperature of 93° F. As can be seen from the table, excellent volatile solids reduction was obtained with good gas production in a relatively short detention time. Not unexpectedly, the solids concentration decreased, the pH increased slightly, and the alkalinity increased by a factor of three.

Gas Production

The current energy crisis has created additional interest in the utilization of sludge gas as an energy source. Many plants for years have used digester gases to heat facilities and drive generators. In general, treatment of 1 mgd of municipal wastewater will provide 1 ton of mixed primary and activated sludge solids which translates to 0.2 to 0.3 lb solids/capita/day. An unheated digester will typically produce 0.32 to 0.56 cu ft of gas/capita while a heated digester will produce from 0.56 to 0.74 cu ft of gas/capita. This is equivalent to a maximum gas production of approximately 11 to 12 cu ft of gas/lb of total solids digested. The heat value of sludge gas is approximately 566 BTU/cu ft. Should electrical generation be considered, approximately 3.5 cu ft of gas is required to produce 1 kilowatt hour (kwh) of electricity [20,21]. Table 5-7 shows the characteristics of sludge gas from several digester installations. As can be seen, the methane content of the gas varies with the feed sludge. Normally, some type of off-gas treatment is necessary to increase the heat content.

TABLE 5-7

CHARACTERISTICS OF SLUDGE GAS [20]

Constituent		**Values for Various Plants**							
		Percent by Volume							
CH_4		42.5	61.0	62.0	67.0	70.0	73.7	75.0	73-75
CO_2		47.7	32.8	38.0	30.0	30.0	17.7	22.0	21-24
H_2		1.7	3.3	trace	–	–	2.1	0.2	1-2
N_2		8.1	2.9	trace	3.0	–	6.5	2.7	1-2
H_2S		–	–	0.15	–	0.01–0.02	0.06	0.1	1-1.5
H_O	BTU/ft^3	459	667	660	624	728	791	716	739-750
d_V	(air = 1)	1.04	0.87	0.92	0.86	0.85	0.74	0.78	0.70-0.80

Supernatant Quality

Typical digester supernatant characteristics appear in Table 5-8. A range of values is given for each parameter since a particular supernatant's quality is dependent upon whether the digester has one or two stages, whether it is mixed, and how well the solids separate from the liquor.

TABLE 5-8

SUPERNATANT CHARACTERISTICS FROM ANAEROBIC DIGESTERS[22]

	Primary Plants (mg/l)	Trickling Filters* (mg/l)	Activated Sludge Plants* (mg/l)
Suspended solids	200-1,000	500- 5,000	5,000-15,000
BOD_5	500-3,000	500- 5,000	1,000-10,000
COD	1,000-5,000	2,000-10,000	3,000-30,000
Ammonia as NH_3	300- 400	400- 600	500- 1,000
Total phosphorus as P	50- 200	100- 300	300- 1,000

*Includes primary sludge.

As can be seen, supernatant from primary sludge digestion requires only minimal concern, when compared to the poor quality supernatant from anaerobically digested mixtures of primary and activated sludges. Methods are available for reducing the amount of activated sludge in the feed to the digester, and this is a way of materially improving the supernatant quality. Methods for treating digester supernatants have been thoroughly discussed in the literature [23]. However, elimination, rather than treatment, of highly polluted digester supernatants is normally sound engineering.

Bacteria Inactivation

Anaerobic digestion reduces bacterial populations as shown in Table 5-9. Although conditions in digesters are unfavorable for the multiplication of most pathogenic organisms, they are not lethal, and the principle bactericidal effect appears to be related to a natural die-off with time.

Impact of Alum and Iron Phosphorus Sludge on Anaerobic Digestion

Chapter 3 indicated that the addition of aluminum sulfate or ferric chloride for improved suspended solids removal and/or phosphorus removal to the primary, biological, or tertiary portion of the treatment plant can substantially increase both the mass and volume of sludge to be treated. Many studies have indicated that neither ferric chloride nor alum phosphorus sludges inhibit anaerobic digestion [24,25,26,27,28]. However, digester performance may be altered due to an increased stress from a higher organic loading and/or lower HRT or feed sludge alkalinity.

TABLE 5-9

BACTERIAL SURVIVAL IN DIGESTION[2]

Bacteria	Digestion Period (days)	Removal (%)	Remarks
Endamoeba hystolytica	12	<100	Greatly reduced populations at 68° F
Salmonella typhosa	20	92	85% reduction in 6 days detention
Tubercle bacilli	35	85	Digestion cannot be relied upon for complete destruction
Escherichia coli	49	<100	Greatly reduced populations at 99° F, about the same reduction in 14 days at 72° F

At Chapel Hill, North Carolina, it was reported [29] that digester alkalinity was reduced in the primary digester from 2,500 mg/l to 1,500 mg/l as $CaCO_3$. This occurred after addition of alum to one of two parallel trains, resulting in a need to add lime on one occasion. Further, the secondary digester underflow concentration decreased from a normal range of 6 to 7 percent to 3.8 percent with a coincident increase in the supernatant SS concentration from 1,000 mg/l to 10,000 mg/l. In spite of these difficulties, the digestion process itself produced a normal reduction in volatile solids throughout the entire alum treatment study.

5.2.5 Upgrading Procedures

The reader is referred to the EPA Technology Transfer Process Design Manual for "Upgrading of Wastewater Treatment Facilities," for a thorough presentation of typical operating results and upgrading procedures [10].

The methods listed here have been successful in improving the efficiency of operating digesters.

- Prethickening of feed sludge.
- Complete mixing of digester.

- Adequate control of operating variables.

- Separate processing of excess air system activated sludge by means other than anaerobic digestion where feasible.

- Where anaerobic digestion of mixtures of primary and excess air system activated sludge is essential, utilize other process modifications to minimize the relative amount of EAS in digester feed and to resolve the solid-liquid separation problem which can result in the second-stage digester.

5.2.6 Typical Costs

Unit costs are presented in Figure 5-7. Use of this cost data should include allowance for the effect of supernatant recycle plus effect of digestion on sludge dewaterability.

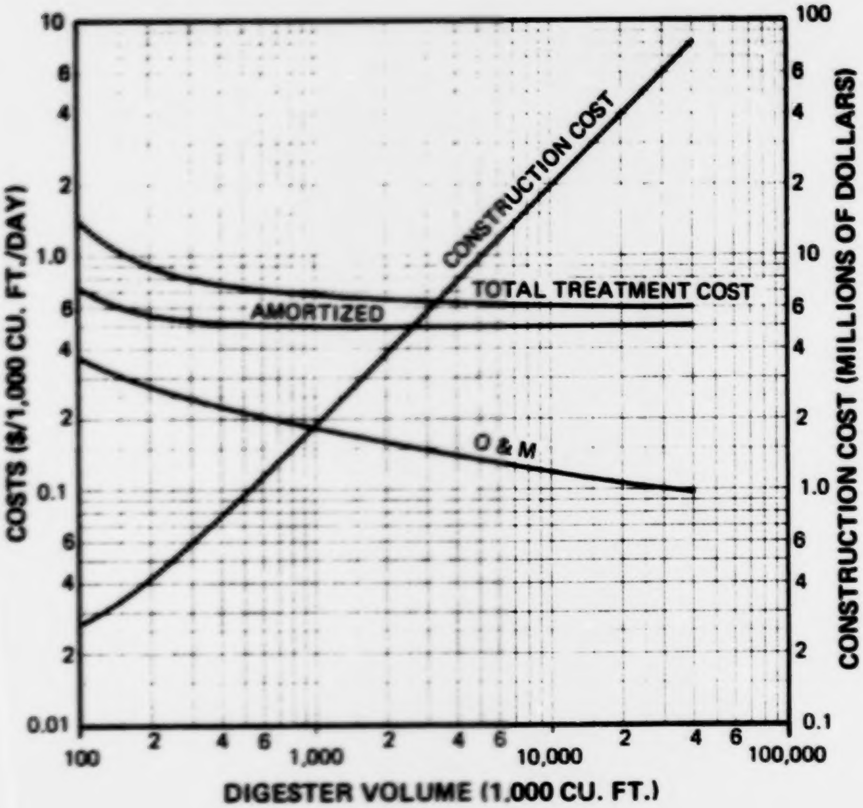
5.3 Aerobic Digestion

Aerobic digestion describes the separate aeration of waste primary sludge, waste biological sludge, or a combination of waste primary and biological sludges in an open tank. It is usually used to stablize excess activated sludges or the excess sludges from small plants which do not have separate primary clarification. Figure 5-8 shows a schematic diagram of an aerobic digestion system. The process involves the direct oxidation of any biodegradable matter by the biologically active mass of organisms and oxidation of microbial cellular material. These two steps are illustrated by the following reactions:

$$\text{organic matter} + O_2 \xrightarrow{\text{Bacteria}} \text{cellular matter} + CO_2 + H_2O$$

$$\text{cellular matter} + O_2 \xrightarrow{\text{Volatile Solid}} \text{digested sludge} + CO_2 + H_2O$$

The second reaction called endogenous respiration is normally the predominant one occurring in aerobic digestion. Stabilization is not complete until there has been an extended period of primarily endogenous respiration (15 to 20 days). Major objectives of aerobic digestion include odor reduction, reduction of biodegradable solids, and improved sludge dewaterability. Process advantages often cited for this process over other stabilization techniques are that it is:



NOTES:

1. Minneapolis, Mar., 1972. ENR Construction Cost Index of 1827.
2. Amortization at 7% for 20 years.
3. Labor rate of $6.25 per hour.
4. Sludge heating, circulating and control equipment and control building included.
5. Source: EPA Cost and Manpower Report and Stanley Consultants.

FIGURE 5-7. Unit anaerobic digestion costs [2].

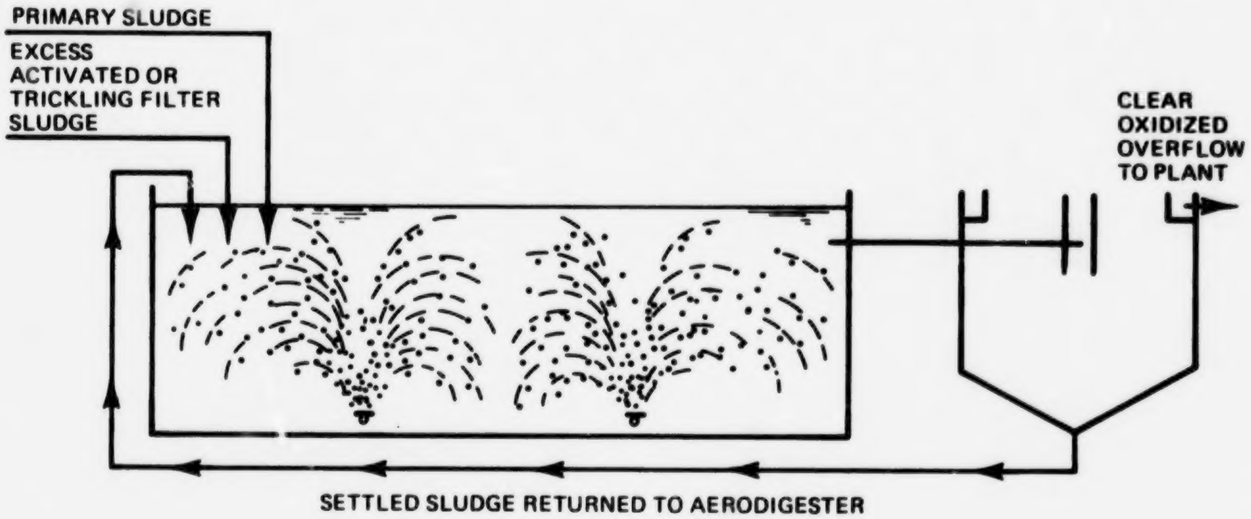

PRIMARY SLUDGE
EXCESS ACTIVATED OR TRICKLING FILTER SLUDGE
CLEAR OXIDIZED OVERFLOW TO PLANT
SETTLED SLUDGE RETURNED TO AERODIGESTER

FIGURE 5-8. Schematic of aerobic digestion system.

- Relatively simple to operate.
- Requires a small capital expenditure compared to anaerobic digestion.
- Does not generate significant odors.
- Reduces the number of pathogenic organisms to a low level.
- Reduces the quantity of grease or hexane solubles.
- Produces a supernatant, if clarified, that is low in BOD, solids, and total P.
- Reduces sludge's respiration rate.

The process is not without disadvantages and these may include high operating cost and unclear design parameters at present. Aerobically stabilized sludge generally has poor dewatering characteristics on vacuum filters [30]. Ordinarily, this sludge is dewatered on sandbeds or applied in liquid form to cropland.

5.3.1 Process Design

The important process parameters are:

- Air or oxygen requirements
- Time of aeration
- Sludge age
- Temperature
- Biodegradable volatile solids
- Processing characteristics of digested sludge
- Supernatant quality

Current practice is to provide approximately 15 days of detention time for the stabilization of excess biological sludges, additional time is required when primary sludge is included. Loadings will normally vary from 0.1 to 0.2 lb VSS/ft^3/day. A 40 to 50 percent reduction in volatile suspended solids content is normally obtained. It is possible for the supernatant to contain as little as 10 to 30 mg/l BOD, 10 mg/l ammonia nitrogen, and from 50 to 100 mg/l nitrate nitrogen [31]. Oxygen requirements, exclusive of nitrification can vary from a minimum of 3 to 30 mg/hr/gm VSS under aeration or higher. When nitrification occurs, both pH and alkalinity are reduced.

Current practice varies according to whether or not a separate sedimentation tank is employed. Smaller plants employ a one-tank, batch-type system, wherein the sludge is supplied with air and completely mixed for a protracted period of time followed by quiescent settling and decantation. Aerobic digesters are designed similarly to rectangular aeration tanks and use conventional aeration systems or they frequently employ circular tanks and use an eductor tube for deep-tank aeration. Figure 5-9 shows a typical circular aerobic digester.

Plants processing significant amounts of sludge frequently incorporate a separate sedimentation basin which facilitates simultaneous decantation and thickening. Aerobic digestion in the past has largely been practiced at small plants on either contact stabilization sludge or on mixtures of primary and activated sludge. Limited plant scale data on the aerobic digestion process is available. Accordingly, wherever possible, pilot work should be conducted prior to design. It should include settling, thickening, and dewatering tests on the aerobically digested sludge alone and in combination with other sludges to be processed. Table 5-10 presents aerobic digestion design parameters with associated remarks [10].

5.3.2 Process Performance Data

Ahlberg and Boyko [33] investigated seven activated sludge facilities treating from 0.03 to 1.40 mgd of municipal wastewater. Some of these facilities had single-stage and others had two-stage digester systems. Table 5-11 summarizes the operation of the facilities investigated. Note the high sludge age and low volatile solids loading of the systems. These sludges should be very well stabilized. In general the sludges fed to the digesters had a pH of 6.7, a COD of 20,000 mg/l, and a total solids concentration of 20,000 mg/l of which 57 percent was volatile. Following digestion, the pH was still around 6.8, the volatile solids were reduced by approximately 50 percent, and the sludge could, on an average, be thickened to about 3 percent solids. Characteristics of the supernatant from settling of the digested sludge are given in Table 5-12. The range of values shown are caused by differences in the solids-liquid separation processes, digester loadings, and nature of the sludges [33].

The Denver, Colorado, District Plant treats 100 mgd of combined domestic and industrial waste by the conventional activated sludge process. Under an Environmental Protection Agency grant, Denver has been investigating aerobic digestion of dilute excess activated sludge on a plant scale [34]. From August 1, 1972, to June 30, 1973, Denver studied, under steady state conditions, the effects of loading rates (0.025 to 0.2 lb VSS/ft^3/day) and the effects of naturally varying temperatures (59 to 84° F) on the efficiency of plant-scale, air-aerobic digestion of excess activated sludge ($<$ 1.0 percent suspended solids concentration). The two most important variables were discovered to be temperature and sludge age. Optimum performance occurred at a loading rate of 0.08 lb VSS/ft^3/day and a temperature of 75° F, which gave a 42 percent VSS destruction. These

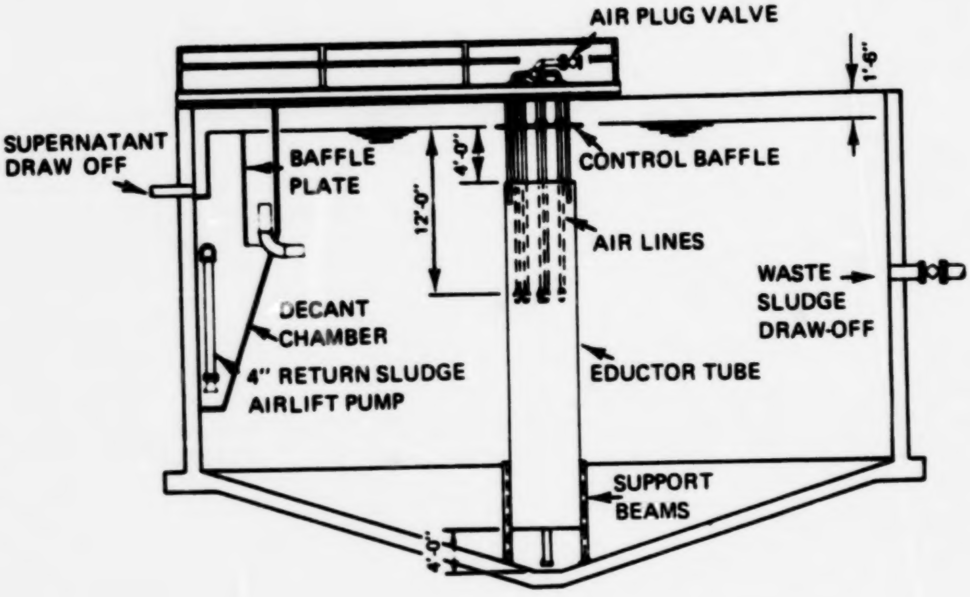

AIR PLUG VALVE
1'-6"
SUPERNATANT DRAW OFF
BAFFLE PLATE
4'-0"
12'-0"
CONTROL BAFFLE
AIR LINES
WASTE SLUDGE DRAW-OFF
DECANT CHAMBER
4" RETURN SLUDGE AIRLIFT PUMP
EDUCTOR TUBE
SUPPORT BEAMS
4'-0"

FIGURE 5-9. Typical circular aerobic digester [32].

TABLE 5-10

AEROBIC DIGESTION DESIGN PARAMETERS

Parameter	Value	Remarks
Solids Retention Time, days	10-15[a]	Depending on temperature, type of sludge, etc.
Solids Retention Time, days	15-20[b]	
Volume Allowance, cu ft/capita	3-4	
VSS Loading, pcf/day	0.024-0.14	Depending on temperature, type of sludge, etc.
Air Requirements		
Diffuser System, cfm/1,000 cu ft	20-35[a]	Enough to keep the solids in suspension and maintain a DO between 1-2 mg/l.
Diffuser System, cfm/1,000 cu ft	>60[b]	
Mechanical System, hp/1,000 cu ft	1.0-1.25	This level is governed by mixing requirements. Most mechanical aerators in aerobic digesters require bottom mixers for solids concentration greater than 8,000 mg/l, especially if deep tanks (>12 feet) are used.
Minimum DO, mg/l	1.0-2.0	
Temperature, ° C	>15	If sludge temperatures are lower than 15° C, additional detention time should be provided so that digestion will occur at the lower biological reaction rates.
VSS Reduction, percent	35-50	
Tank Design		Aerobic digestion tanks are open and generally require no special heat transfer equipment or insulation. For small treatment systems (0.1 mgd), the tank design should be flexible enough so that the digester tank can also act as a sludge thickening unit. If thickening is to be utilized in the aeration tank, sock type diffusers should be used to minimize clogging.
Power Requirement, BHP/10,000 Population Equivalent	8-10	

[a] Excess activated sludge alone.

[b] Primary and excess activated sludge, or primary sludge alone.

TABLE 5-11

SUMMARY OF AEROBIC DIGESTION OPERATION

Parameter	Range
HRT (days)	14-360
Sludge Age (days)	29-320
Actual Loading (lb VS/cu ft/day)	0.0035-0.027
Air Supply (cfm/1,000 cu ft)	8.4-46

TABLE 5-12

CHARACTERISTICS OF AEROBIC DIGESTION SUPERNATANT

Parameter	Overall Average	Range
pH	7.0	5.9-7.7
	mg/l	
BOD_5	500	9-1,700
Filtered BOD_5	51	4-183
COD	2,600	288-8,140
SS	3,400	46-11,500
Kjeldahl N	170	10-400
Total P	98	19-241
Soluble P	26	2.5-64.0

conditions also coincided with an oxygen requirement of 1.2 to 1.6 lb O_2/lb VSS, a dissolved oxygen level of 1 to 2 mg/l in the mixed liquor, a hydraulic detention time of about 5 days, and an aeration rate of 30 cfm/1,000 ft^3. The Denver project is discussed in more detail as a case history in Chapter 10.

5.3.3 Oxygen Aerobic Digestion

Pure oxygen, rather than air, may be used in aerobic digestion to stabilize thicker sludges in which the high oxygen uptake rates cannot be satisfied with air aeration. A three-month plant scale study of aerobic digestion using pure oxygen in a closed system was conducted at Speedway, Indiana [35]. Oxygenation took place in a covered 31,000 cu ft, four-stage reactor followed by a clarifier for decanting the sludge. The study was divided into two phases, the first treating only excess activated sludge and the second treating mixed primary and excess activated sludge. Much of the heat generated by biological oxidation was retained within the closed system. This resulted in a significant increase in sludge temperature and a corresponding increase in the rate of VSS destruction. The results of this study are shown in Table 5-13. A positive DO level of 2 mg/l was maintained.

TABLE 5-13

RESULTS OF HIGH-PURITY OXYGEN AEROBIC DIGESTERS
SPEEDWAY, INDIANA

Parameter	Phase 1	Phase 2
Feed Sludge Type	EAS[a]	57% EAS + 43% PS[b]
Feed TSS, percent	2.14	3.06
VSS/TSS	0.17	0.66
Biodegradable VSS/TSS	0.45	0.48
Feed Temperature, °C	19.5	16.0
Average Ambient Air Temperature, °C	7.5	-2.2
Sludge Temperature–Stage 4, °C	33.0	31.0
HRT, days	16.3	11.6
Volatile Solids Loading Rate, lb/ft^3/day	0.064	0.109
VSS Reduction, percent	44	43
Biodegradable VSS Reduction, percent	94.6	86.6

[a] Excess activated sludge.

[b] Primary sludge.

5.3.4 Aerobic Digestion Costs

Ritter [36], in 1970, estimated the average power costs of aerobic digestion for three small communities in Pennsylvania, at about $2.18/yr/lb BOD received per day, or $0.37/yr/capita served. Capital cost for aerobic digestion is estimated in Figure 5-10. Evaluation of aerobic digestion versus anaerobic digestion should take into account the effect of unit process selection on overall system cost as well as the operating and capital costs of the two-unit processes.

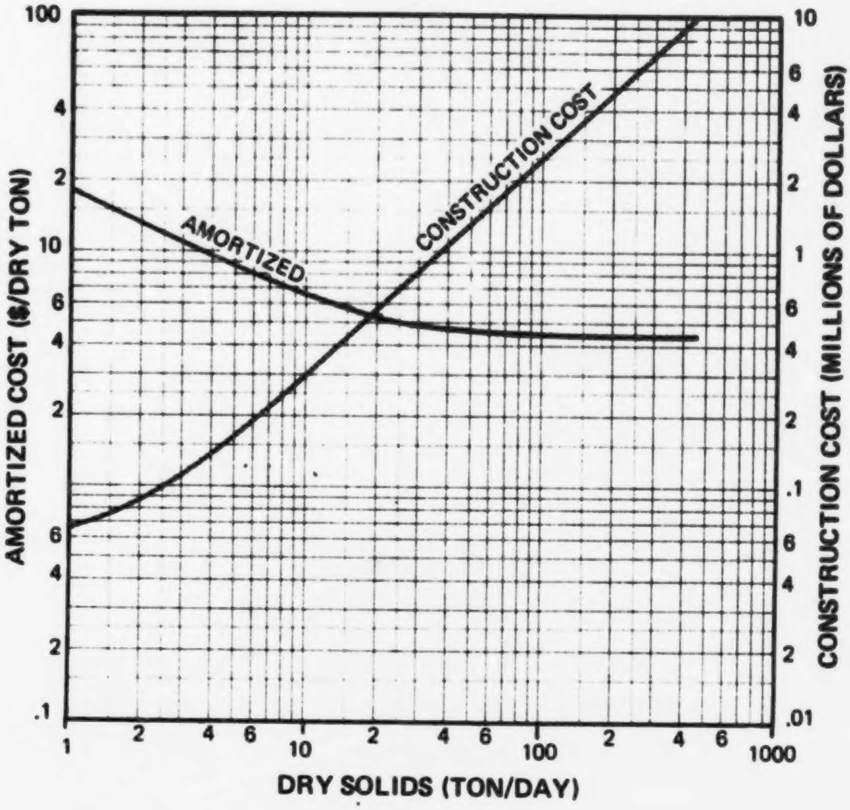
5.4 Chlorine Oxidation

The Purifax process oxidizes sludge with heavy doses of chlorine (about 2,000 mg/l). Following treatment, the sludge dewaters well on sandbeds and it is stable. Purifaxed sludges may require chemical conditioning prior to dewatering on vacuum filters, since the sludge after treatment has a low pH (about 2). Supernatants and filtrates from the process contain high concentrations of chloramines. A first estimate of cost would be approximately $5/ton of dry sludge solids for the purchase of chlorine only. Other operating costs and capital costs would increase this figure.

5.5 Lime Treatment

The addition of lime, in sufficient quantities to maintain a high pH between 11.0 and 11.5, stablizes sludge and destroys pathogenic bacteria. Lime stablized sludges dewater well on sandbeds without odor problems. Sludge filterability can be improved with the use of lime; however, caution is required when sludge cake disposal to land is practiced. Disposal in thick layers could create a situation where the pH could fall to near 7 prior to the sludge drying out, causing regrowth of organisms and resulting noxious conditions. Essentially, no organic destruction occurs with lime treatment. The key factor in assuring a proper stabilization process is the maintenance of a pH of around 11.0.

Farrell et al. [37] recently investigated the lime stabilization of sludge at the 1.15 mgd Lebanon, Ohio, wastewater treatment plant. While this plant had an anaerobic digester, it needed a simple, reliable, and inexpensive sludge treatment process to handle the excessive solids produced by the upgrading of the facility for phosphorus removal. Both iron and aluminum additions to the primary portion of the plant were employed. The procedure followed in this attempt to achieve lime stabilization included the addition of sufficient lime [$Ca(OH)_2$] to elevate the sludge to a pH of 11.5, where it remained for 30 minutes. The dosage essentially maintained the pH above 11.0 for 24 hours. The mixing of the lime slurry and the sludge was accomplished using air mixing.



NOTES:

1. Minneapolis. Mar., 1972. ENR Construction Cost Index of 1827.
2. Amortization at 7% for 20 years.
3. Influent sludge of 38% primary and 62% waste activated sludge with a solids content of 3.5%.
4. 20 day volumetric displacement time.
5. Souce: EPA Cost and Manpower Report and Stanley Consultants.

FIGURE 5-10. Aerobic digestion capital cost [2].

The effect of lime treatment on typical pathogenic bacteria present in sludge is shown in Table 5-14. As can be seen, salmonella and pseudomonas were totally eliminated and the total aerobic count was reduced by between 88 and greater than 99 percent. In later testing, the fecal coliform and fecal streptococci were shown to be destroyed by greater than 99 percent.

TABLE 5-14

BACTERIOLOGICAL STUDIES OF SLUDGE PRODUCED IN PLANT-SCALE TESTS AT LEBANON

Sludge	Bacterial Count (organisms/l of sludge) Salmonella Species	Pseudomonas aeruginosa	Total Aerobic Count × 10^{-8}
Alum-primary	110	1,300	41
Limed alum-primary	None detected	None detected	5.0
Ferric-primary	>24,000	610	190
Limed ferric-primary	None detected	None detected	0.29

The effect of lime stabilization on the vacuum filterability of the sludge was studied in the laboratory with a filter leaf apparatus. The results are shown in Table 5-15. Note the filter yield in every case was increased by a factor of about two, while cake moisture was essentially unchanged.

TABLE 5-15

EFFECT OF LIME ON FILTERABILITY OF ALUMINUM AND IRON PRIMARY SLUDGES AT LEBANON

Sludge Property	Lime Addition	Al^{+++} Dose (mg/l)			Fe^{+++} Dose (mg/l)	
Filter leaf test yield	Before	0.98	0.94	0.95	1.06	1.57
(lb/hr/ft^2)	After	1.97	2.10	2.58	1.57	2.40
Cake moisture	Before	4.31	4.35	4.37	4.10	3.75
(lb water/lb dry solids)	After	3.87	3.92	3.83	4.28	3.75

In every instance, the lime conditioned sludge exhibited no obnoxious odors and because high pH conditions were present, the ammonia nitrogen concentration was reduced by approximately 50 percent as a result of air stripping. The average cost for lime addition is shown in Table 5-16. This information is based on a hydrated lime cost of $20/ton.

TABLE 5-16

AVERAGE COST OF LIME ADDITION–PLANT-SCALE TESTS AT LEBANON

Treatment of Raw Wastewater		Resultant Sludge Solids	$Ca(OH)_2$ Added to Sludge		Lime Cost
Chemical	Dose (mg/l)	(g/l)	(g/l)	(lb/ton)	($/ton)
Al^{+++}	31.8	24.0	6.0	500	5.0
	22.7	28.5	6.3	440	4.4
	13.6	20.6	4.4	420	4.2
Fe^{+++}	31.0	20.2	2.2	220	2.2
	15.5	17.9	2.2	240	2.4

More recent information has been obtained through an EPA–Battelle study [38] on the evaluation of lime stabilization. Results show that the pH must be maintained between 12.2 and 12.4 to insure that the sludge is stabilized and the pH should remain above 11.0 for better than 2 weeks. Paulsrud and Eikum [39] agree with these findings and determined the lime doses required to keep the sludge at a pH greater than 11.0 for 14 days. This information is shown in Table 5-17. Battelle has estimated the costs for lime addition to a pH of 12.2 to 12.4 including all operation and maintenance costs to range from $8/ton for primary sludges to approximately $15/ton for biological sludges.

TABLE 5-17

LIME DOSE REQUIRED TO KEEP SLUDGE pH >11.0 FOR AT LEAST 14 DAYS

Type	Dose (lb $Ca(OH)_2$ /ton sludge solids)
Primary sludge	200- 300
Septic tank sludge	200- 600
Biological sludge	600-1,000
Al sludge (secondary precipitation)	800-1,200
Al sludge (secondary precipitation) + Primary sludge ($SS_{Al} : SS_{Prim} = 1:1$)	500- 800
Fe sludge (secondary precipitation)	700-1,200

5.6 Heat Treatment for Stabilization

Heat treatment is a well known method of destroying pathogenic organisms and has been applied successfully for disinfecting sludge. Two methods that have been applied to sludge treatment are pasteurization and low pressure oxidation.

Pasteurization implies heating to a specific temperature for a time period that will destory undesirable organisms in sludge. In West Germany and Switzerland, pasteurization is required when sludge is spread on pastures during summer growth periods. Table 5-18 shows the effect of various pasteurization temperatures and times on typical pathogenic organisms found in sludge [40]. Stern [41] recently concluded that pasteurization at 70° C for 30 to 60 minutes is effective for destroying pathogens in digested sludge. About 75° C for one hour is effective in reducing coliform indicators below 1,000 counts per 100 ml as well as destroying pathogens. Stern's estimated cost for pasteurization ranged from $5.00 to $22.00 per ton of dry sludge solids and was dependent upon the size plant, the fuel source, and whether heat recuperation was employed.

TABLE 5-18

EFFECT OF TIME AND TEMPERATURE ON THE SURVIVAL OF TYPICAL PATHOGENS FOUND IN SLUDGE[*]

Organism	Temperature ° C				
	50	55	60	65	70
	minutes				
Cysts of *Entamoeba histolytica*	5				
Eggs of *Ascaris lumbricoides*	60	7			
Brucella abortus		60		3	
Corynebacterium diphtheria		45			4
Salmonella typhosa			30		4
Escherichia coli			60		5
Micrococcus pyrogene var. *aursus*					20
Mycobacterium tuberculosis var. *promixis*					20
Viruses					25

[*] Pathogens completely eliminated at indicated time and temperature.

Under normal conditions of low pressure oxidation (LPO) employment, which is to condition the sludge for improved dewaterability, all pathogenic organisms are destroyed due to the high temperatures achieved and the retention time. Typically, the sludge temperature is elevated to between 350° and 400° F, pressure is raised to 180 to 210 psi, and the retention time is between 20 and 30 minutes. Chapters 6 and 11 discuss the LPO systems in more detail.

5.7 Composting

Composting of sludge, either separately or with municipal solid wastes has not been widely applied in North America. Of the 18 plants constructed in the United States between 1951 and 1969, few are currently operated and many of these are operated only intermittently. The primary problem has been lack of a market for the stable product and a market is required to produce revenue from the product's sale. This revenue offsets the cost of the process and can make it economical. Many of the processes involve composting in windrows with mechanical turning to provide oxygen for the microorganisms to carry out the stabilization process.

5.7.1 Process Description

Though there are over 30 composting systems identified by inventor or proprietary name, in general, the methods can be broadly classified by the digestion procedure employed. Digestion is accomplished by the windrow with intermittent mixing procedure or by aeration in a mechanical device.

Sequential steps usually involved in composting are:

- *Preparation* Sludge that is composted without the inclusion of solid waste fractions, must be blended with some bulking material if windrows are to be used. This bulking material can be soil, sawdust, wood chips, etc. If a mechanical aeration system is used: bulking agent requirements are less severe. For good digestion of the waste, a moisture content between 45 and 65 percent by wet weight is desirable. A potential advantage exists when combined sludge solid waste composting is practiced, because the digested sludge can provide nutrients and requisite moisture to the solid waste fraction. Normal sludge to refuse ratios of roughly 0.50 to 1.00 by weight are employed.

- *Digestion* The digestion period is characterized by rapid decomposition. Air is supplied by periodic turnings in windrow-type operations. While in mechanical systems, forced draft or agitation in long screw conveyors is utilized. The reaction is exothermic and the wastes reach temperatures of 140° F to 160° F or higher. Pathogen kill and the inactivation of objectionable materials are possible at these temperatures. Period of digestion is normally about six weeks for windrows and several days in mechanical aeration systems. Although the ratio of carbon to nitrogen (C/N) can be used as a rough estimate of completion, careful testing is required to determine operable conditions. A C/N ratio of 20 has been suggested as a theoretical upper limit. This ratio prevents robbing of nitrogen from soil treated with compost by further microbial activity.

- *Curing*–This is characterized by a slowing of the decomposition rate. The temperature drops back to normal and the process is brought to completion. This period takes about two more windrow weeks or one to two weeks for mechanical systems.

- *Finishing*–If municipal solid waste fractions containing nondigestible debris have been included, some sort of screening or other removal procedure may be necessary.

5.7.2 Process Performance and Costs

Time and temperature are two major design considerations since they effect the hygienic quality of the composting product as shown in Table 5-19 [42]. From this table it can be seen that the water content of the material entering the system ranges between 40 and 60 percent. Typically, a mechanical system can achieve a temperature in excess of 60° C and frequently in excess of 65° C. These materials are then pathogen-free in as little time as one day and free of all spore formers in approximately one week.

Volume and weight reductions hinge on the type of mixture fed to the composting process, so generalizations are not of great value. It has however been observed that from 20 to 50 percent less space is required in a landfill when the waste material has been composted. Composting costs vary widely and ranges of $2 to $20 per ton have been reported. Some recent capital and operating costs are shown in Figure 5-11. Revenue from the sale of the composted product is usually $1.50 to $3.50 per ton of material fed.

5.8 Additional Reading

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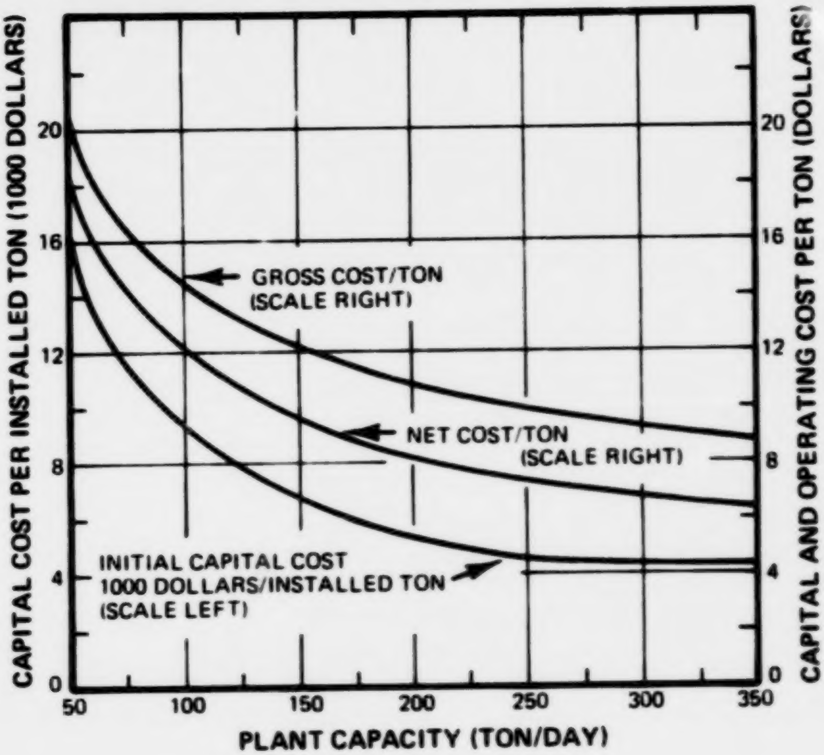
TABLE 5-19

HYGIENIC QUALITY OF COMPOST

Treatment Method	Material	Water Content (%)	Maximum Temp. Achieved (°C)	Hygienic Evaluation	Remarks
Contour Composting					
• Contour spreading	sludge + solid waste	55	46	Not pathogen-free after 5 months	-
• Windrow spreading	sludge	60	52	Not pathogen-free after 6 months	-
• Windrow spreading	solid waste	40-60	>55	Pathogen-free after 3 weeks	-
• Windrow spreading	sludge + solid waste	40-60	>55	Pathogen-free after 3 weeks	-
Mechanical Composting					
• Rotating drum (Dano Process)	solid waste	45-55	>60	Pathogen-free after 6-7 days	• Spore-free after 1 week of windrow composting
• Rotating drum	sludge + solid waste	approx 50	>60	Pathogen-free after 6-7 days	• Spore-free after 1 week of windrow composting
• Rotating tower (Multibacto process)	solid waste	40-50	>65	Pathogen-free after 1 day	• Spore-free after 1 week of windrow composting
• Rotating tower	sludge + solid waste	45-55	>65	Pathogen-free after 1 day	• Spore-free after 1 week of windrow composting



NOTES:

1. Plant capacity is normally one or two shifts per day to achieve plant capacity.
2. Gross cost trend is the owning and operating facilities without any credits.
3. Net cost trend is for owning and operating facilities considering sales of compost and salvaged materials.
4. All costs consider compost digested sludge with refuse.
5. Source: Composting of Municipal Solid Wastes in the United States, US Environmental Protection Agency (1971).

FIGURE 5-11. Composting costs [2].

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CHAPTER 6

SLUDGE CONDITIONING

6.1 Functions, Methods, and Occurrences

Conditioning is herein defined as the pretreatment of sludge to facilitate water removal by a thickening or dewatering process. Table 6-1 shows the usual conditioning methods, the unit processes they are employed with, and the purposes they serve.

TABLE 6-1

CONDITIONING METHODS AND PURPOSES

Conditioning Method	Unit Process	Function
Polymer Addition	Thickening	Improve loading rate, degree of concentration, and solids capture.
Polymer Addition	Dewatering	Improve production rate, cake solids content, and solids capture.
Inorganic Chemical Addition	Dewatering	Improve production rate, cake solids content, and solids capture.
Elutriation	Dewatering	Decrease acidic chemical conditioner demand and increase degree of concentration.
Heat Treatment	Dewatering	Eliminate or decrease chemical use, improve production rate, cake solids content, and stabilization. Some conversion may also occur.
Ash Addition	Dewatering	Provide improved cake release from belt type vacuum filters and facilitate filter pressing. It can also result in higher filter yields and reduced chemical requirements.

It should be noted that the first three methods involve addition of chemicals to coagulate and/or flocculate the sludge to accomplish the functions listed. The methods do not otherwise materially change the nature of the sludges. The inorganic chemicals, however, can change the solution pH, increase the inorganic fraction of sludge, and also affect stabilization. Ferric chloride addition with or without lime has been the principal method for conditioning sludges prior to dewatering. In the past ten to fifteen years, however, polymer conditioning before dewatering has become widespread. The metal salts are generally only used now where polymers have not yet been demonstrated as economically effective or on sludges where polymers will not work. Polymers are also widely used for flotation and centrifugal thickening and they are occasionally used in gravity thickening.

Elutriation was originally developed to decrease the alkalinity of anaerobically digested sludges which reduced the demand for acidic metal salts. It is not now as widely practiced as in previous years, although elutriation basins are still employed in many plants for postanaerobic digestion thickening. Elutriation, depending upon the particular system's design and operation and the type of sludge being treated, can result in fractionation of the sludge by particle size and density. This fractionation can result in a serious recirculation side-stream effect.

Heat treatment facilitates dewatering. At the same time it solubilizes a portion of the treated sludge. Depending on the type of sludge treated and the kind of system employed for treatment, this solubilization of sludge may create a cooking liquor recirculation stream which requires a separate treatment system. This is particularly true of process sludges which contain a high percentage of activated sludge. Naturally, treatment of this recirculation stream will generate additional sludge to be handled. Heat treatment also kills pathogenic organisms. Heat treatment was originally practiced in Great Britain in the 1940s as a means of processing difficult sludges. As of 1971 about 11 percent of the British population were served by plants employing heat treatment. Heat treatment systems are currently being designed into a number of U.S. plants [1].

Ash addition to sludge for improving the dewatering operation is occasionally employed. Sludge incinerator ash is added to the sludge at Indianapolis and Cedar Rapids to improve cake release and reduce the chemical requirements for good filter productivity and cake solids concentration.

6.2 Considerations in Selecting a Conditioning Method

In the past few years, the impact of sludge processing systems on total wastewater plant capital and operating costs has been considered more thoroughly. Sludge thickening and dewatering can materially affect the preceding and succeeding unit processes. The conditioning method chosen normally has a significant effect on the efficiency of the thickening and/or dewatering operation. It should also be realized that the method of conditioning can have a pronounced effect on the liquid treatment portion of the plant. This can best be illustrated by a hypothetical example. Consider heat treatment, although the same kind of example could be applied to other conditioning processes, for the conditioning of combined primary and excess activated sludge from a conventional activated sludge plant as shown in Figure 6-1.

The proposed plant will treat 100 mgd of raw municipal wastewater containing 187,000 lb of BOD_5 and 98,000 lb of suspended solids. The primary portion of the plant is expected to remove 50 percent of the suspended solids and 25 percent of the BOD_5. The primary sludge will have a solids concentration of 3 percent which can be raised to 6 percent by gravity thickening. The activated sludge portion of the plant will operate with a MLSS of 3,000 mg/l and a detention time of 6.5 hr. The secondary portion is expected to remove 81 percent of the BOD_5 remaining after primary clarification. It was estimated that 1.065 lb of EAS would result from removal of 1.0 lb of BOD_5 in the aeration system and these solids can be air flotation thickened to 4 percent. These data are summarized in Table 6-2.

TABLE 6-2

PLANT DESIGN CRITERIA

Primary Treatment Sludges	
Influent suspended solids, lb/day	98,000
Primary sludge solids to gravity thickener, lb/day	49,000
Primary sludge solids concentration, percent	3.0
Thickened sludge solids to hold tank, lb/day	47,000
Thickened sludge solids concentration, percent	6.0
Aeration–Final Clarifier System	
MLSS mg/l	3,000
Influent BOD_5, lb/day	140,000
BOD_5 removed, lb/day	113,700
Detention time, hr	6.5
Flotation Thickener	
Influent excess activated sludge solids, lb/day	121,000
Thickened sludge solids to hold tank, lb/day	109,000
Thickened sludge solids concentration, percent	4.0

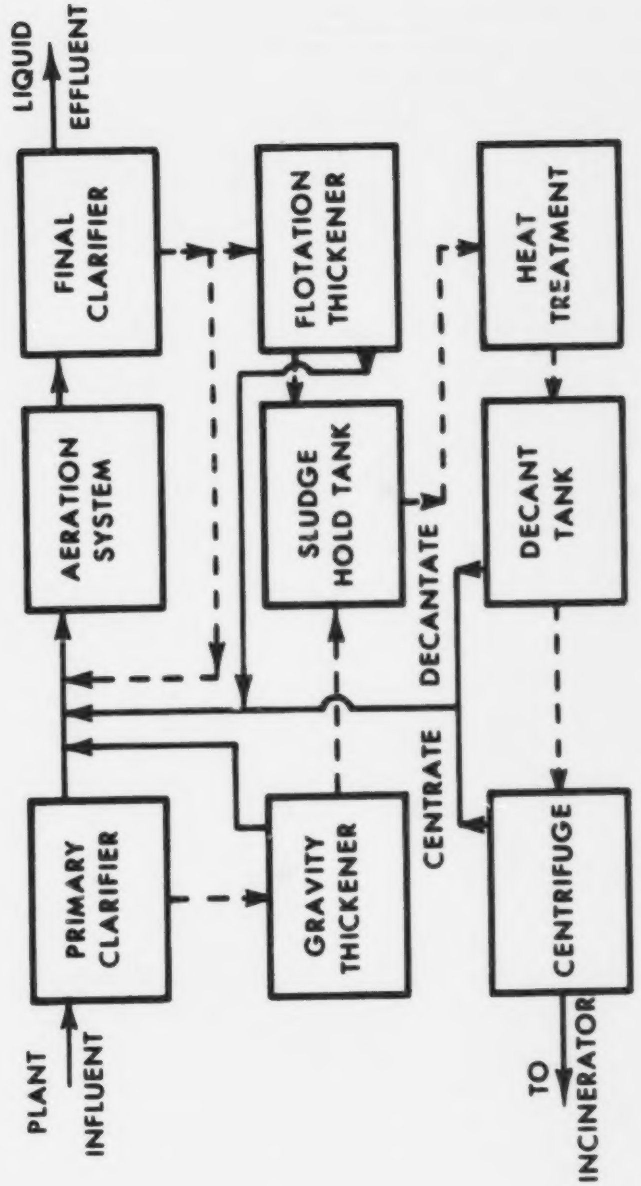

PLANT INFLUENT
PRIMARY CLARIFIER
AERATION SYSTEM
FINAL CLARIFIER
LIQUID EFFLUENT
GRAVITY THICKENER
SLUDGE HOLD TANK
FLOTATION THICKENER
CENTRATE
DECANTATE
CENTRIFUGE
DECANT TANK
HEAT TREATMENT
TO INCINERATOR

FIGURE 6-1. Conceptual flow sheet wastewater plant with heat treatment.

In the proposed design, primary sludge would be thickened by gravity, and the EAS would be air flotation thickened. After combining the two thickened sludges in a holding tank, they would be heat treated, decanted, centrifuged, and incinerated. Assumed solids capture for the various unit processes are primary clarification, 50 percent; gravity thickener, 96 percent; DAF, 90 percent; heat treatment decant tank, 95 percent; and the dewatering centrifuge, 90 percent. During heat treatment, 20 percent of the solids are destroyed by oxidation. The quantities of solids entering and leaving the heat treatment, thickening, and dewatering unit processes together with a summary of the solids loss during processing are given in Table 6-3. This table clearly shows the effect of solids recirculation on the aeration portion of the plant. In this instance recirculation from the various solids processing systems has increased the load on the aerator by 65 percent.

TABLE 6-3

ADDITIONAL DESIGN CRITERIA

Combined Sludge Processing

Heat Treatment	
Influent sludge solids, lb/day	156,000
Influent sludge solids concentration, percent	4.45
Sludge solids oxidized, lb/day	31,000
Thickening (Decant Tank)	
Influent sludge solids, lb/day	125,000
Influent sludge concentration, percent	3.6
Dewatering (Centrifuge)	
Influent sludge solids, lb/day	119,000
Influent sludge solids, percent	8.0
Dewatered sludge solids, lb/day	107,000
Dewatered sludge solids concentration, percent	30

Summary of Sludge Solids to Aeration System

	% solids lost	lb/day
Recirculated from sludge processing		
Gravity thickener overflow	4	2,000
Flotation thickener underflow	10	12,000
Heat treatment decantate	5	6,000
Heat treatment centrate	10	12,000
Subtotal		32,000
Primary clarifier effluent	50	49,000
Total		**81,000**

Casual examination indicates that the illustrated concept of Figure 6-1 and the preceding design criteria could be effective with allowances for additional solids handling. Closer study reveals that consideration of dissolved solids loadings generated in heat treatment and recirculated was neglected. An unknown is the degree of sludge solubilization during heat treatment. Material balances presented in Table 6-4 were prepared to explore this unknown. Case I in this figure represents the concept described above. Various other degrees of solubilization are assumed in Cases II and III for comparison. Table 6-5 presents a summary of solids loading recirculated from heat treatment, as derived from Table 6-4.

The concept, described above, provided for processing of 156,000 lb/day of sludge solids. It allowed for recirculation of 18,000 lb/day of suspended solids from heat treatment. However, it now appears that the solids recirculated from this source could be considerable and include a significant amount of solubilized solids. Imposition of the additional recirculated solids requires consideration in the conceptual design.

Organic, suspended, and dissolved solids can generate excess activated sludge when placed under aeration. Substantial loadings of recirculated solids would, therefore, be expected to have an effect on the treatment plant's efficiency. A plant confronted with this type of situation may have to expand the liquid treatment portion to accommodate the recirculation load.

The above example illustrates the importance of sludge conditioning process evaluation and selection in overall plant design. While this particular example illustrated a system involving heat treatment, later case studies will discuss other conditioning systems. The crux of the problem is to minimize recirculation effects in an economic fashion.

6.3 Process Chemistry–Conditioning

Each of the various sludge conditioning methods functions in a different way and causes diverse chemical and physical effects. Conditioning by either organic or inorganic chemical addition, elutriation, and heat treatment are discussed.

6.3.1 Chemical Conditioning and the Use of Polyelectrolytes

The solid particles present in sludge usually necessitate conditioning because they are fine in particle size, hydrated, and carry an electrostatic charge. These particle properties inhibit thickening and dewatering and are caused by the chemical composition and surface structure of the particles. Figure 6-2 depicts a typical liquid-solid interface relationship, and it helps to explain the difficulty so often encountered when releasing water from sludge. Sludge in effect is a stable colloidal suspension and it is the function of a conditioner to destabilize the suspension. The particles can lose, gain, or share electrons by forming covalent, ionic, hydrogen, dipolar, or induced dipolar bonds.

TABLE 6-4

COMPARATIVE SOLIDS BALANCES
VARIOUS SLUDGE PROCESSING CONDITIONS

	Case					
	I		II		III	
Assumed conditions						
Influent sludge dissolved solids	Negligible		Negligible		Negligible	
			%			
Heat treatment process						
Suspended solids oxidized	20		20		None	
Suspended solids solubilized	None		20		40	
Thickening (decantation) process						
Suspended solids capture	95		95		95	
Thickened sludge solids content	8		8		8	
Dewatering (centrifugal) process						
Suspended solids capture	90		90		90	
Cake solids content	30		30		30	
			lb/day			
Influent sludge suspended solids (Solids concentration, 4.45%)	156,000 ↓		156,000 ↓		156,000 ↓	
	Heat Treatment		Heat Treatment		Heat Treatment	
	↓ Oxidized		↓ Oxidized			
Solubilized solids		None		31,000		62,000
Suspended solids	31,000	125,000	31,000	94,000		94,000
Total solids		125,000 ↓		125,000 ↓		156,000 ↓
	Thickening		Thickening		Thickening	
	↓ Decantate		↓ Decantate		↓ Decantate	
Solubilized solids	None	None	20,400	10,600	37,900	24,100
Suspended solids	6,000	119,000	4,700	89,300	4,700	89,300
Total solids	6,000	119,000 ↓	25,100	99,100 ↓	42,600	113,400 ↓
	Dewatering		Dewatering		Dewatering	
	↓ Centrate	↓	↓ Centrate	↓	↓ Centrate	↓
Solubilized solids	None	None	8,800	1,800	20,500	3,600
Suspended solids	12,000	107,000	8,900	80,400	8,900	80,400
Total solids	12,000	107,000	17,700	82,200	29,400	84,000

Note: Processing by heat treatment, thickening, and dewatering as shown in Figure 6-1.

TABLE 6-5

RECIRCULATED SOLIDS LOADINGS DURING DEWATERING

Type of Solids	Case I	 II	 III
	lb/day		
Solubilized	None	29,200	58,400
Suspended	18,000	13,600	13,600
Total	18,000	42,800	72,000
	%		
Total Solids concentration*	0.58	1.34	2.23

*In combined centrate and decantate.

Practically all dispersed particles in wastewater carry a net negative charge. Depending on the net charge of the suspension it may be flocculated by adsorption of polymers (polyelectrolytes) with either positive (cationic) or negative (anionic) charges. Nonionic (net zero charge) polymers also may function as flocculants. Anionic polymers adsorb via isolated cationic sites on the particles. Nonionic polymers can adsorb through hydrogen bonding, but they also normally assume some degree of negative Zeta potential in dispersed form.

In general, the polyelectrolytes flocculate by neutralizing the surface charges on the dispersed particles by causing the desorption of bound water and through bridging. The latter is simultaneous attachment of the polymer to two or more solid particles. Figure 6-3 shows the variety of configurations that are thought to be involved in the flocculation mechanism. The desorption of bound water, neutralization of surface charges, and aggregation which are brought about by flocculants result in the formation of structured free-draining cakes during dewatering operations. In gravity thickening the aggregates settle more rapidly to a higher solids content, while in flotation thickening aggregation of particles and greater compaction occur with the use of polyelectrolytes. The polymeric flocculants most useful in conditioning are largely linear, high molecular weight materials which carry a large charge density in an aqueous dispersion form. Figure 6-4 shows the chemical composition of the repeating monomeric units for a typical polyanionic and polycationic conditioner. The molecular weight of useful materials is of the order of 200,000 to 10 million. Figure 6-5 shows the typical configuration of a polyelectrolyte in solution. This simplified figure does not show the tremendous length of the polymer's molecule chain.

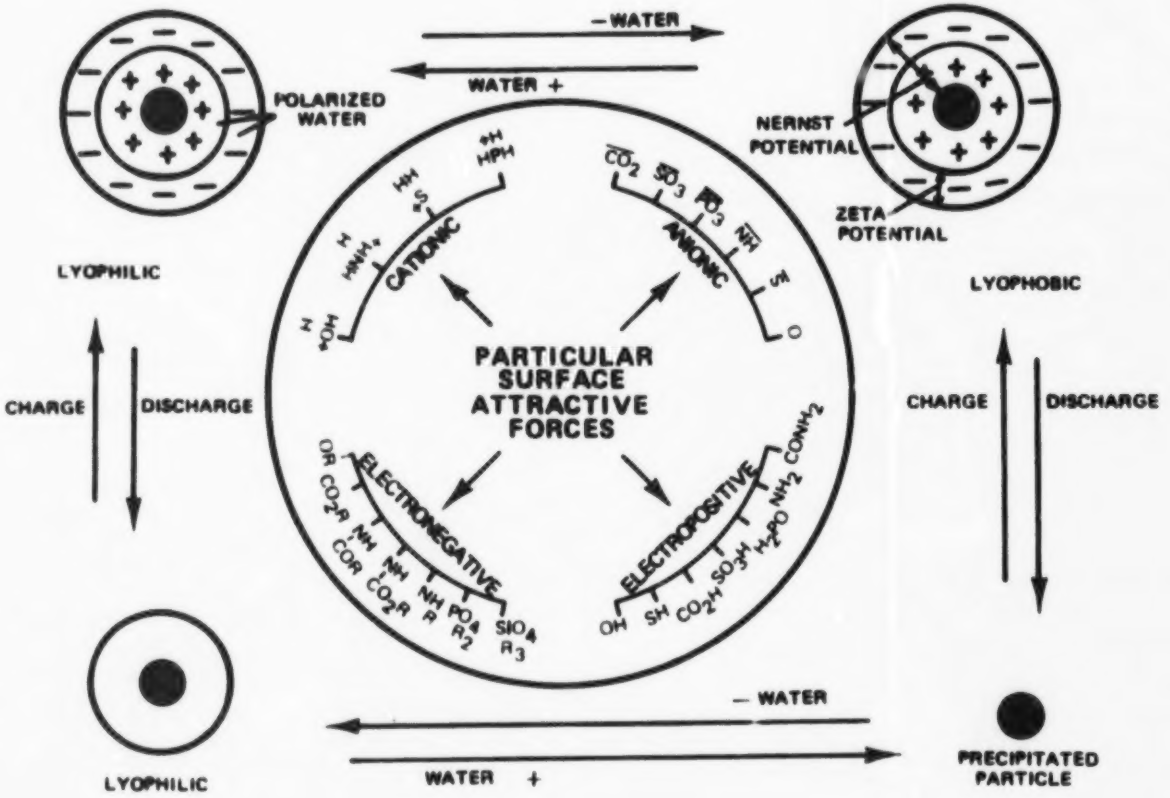
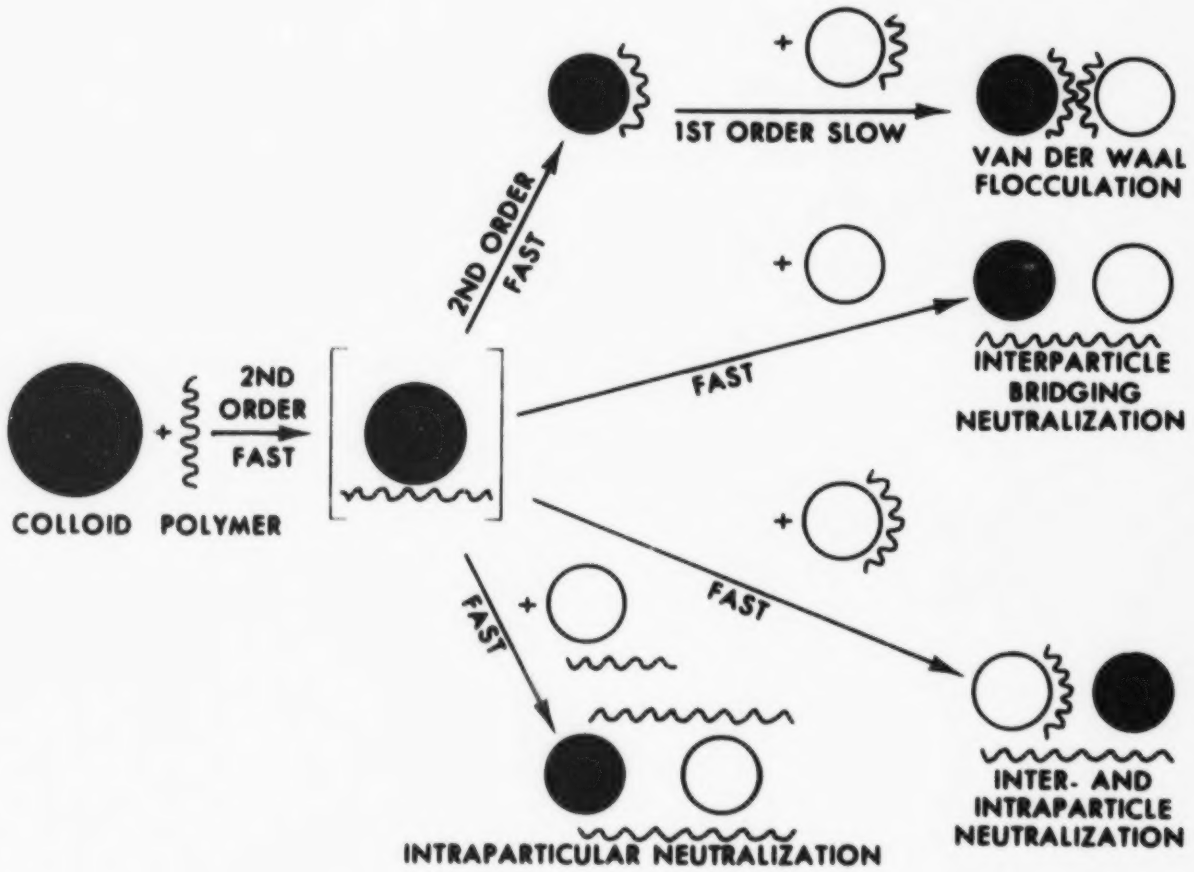
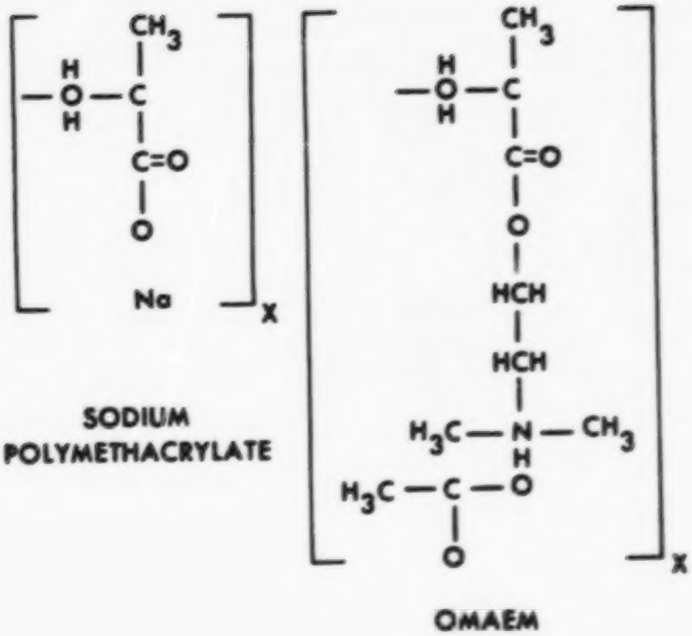

-WATER
WATER +
POLARIZED WATER
NERNST POTENTIAL
ZETA POTENTIAL
LYOPHILIC
LYOPHOBIC
CHARGE
DISCHARGE
CHARGE
DISCHARGE
PARTICULAR SURFACE ATTRACTIVE FORCES
CATIONIC
ANIONIC
ELECTRONEGATIVE
ELECTROPOSITIVE
CO2 SO3 PO3 NH S O
OH SH CO2H SO3H H2PO NH2 CONH2
OR CO2R NH COR NH CO2R NH R PO4 R2 SIO4 R3
— WATER
WATER +
LYOPHILIC
PRECIPITATED PARTICLE

FIGURE 6-2. Factors influencing the stability of a colloidal suspension [2].



2ND ORDER FAST
1ST ORDER SLOW
VAN DER WAAL FLOCCULATION
2ND ORDER FAST
FAST
INTERPARTICLE BRIDGING NEUTRALIZATION
COLLOID
POLYMER
FAST
INTER- AND INTRAPARTICLE NEUTRALIZATION
FAST
INTRAPARTICULAR NEUTRALIZATION

FIGURE 6-3. Mechanism of polymer flocculation [2].



SODIUM POLYMETHACRYLATE

OMAEM

FIGURE 6-4. Structure of two polyelectrolytes's monomeric units [3].

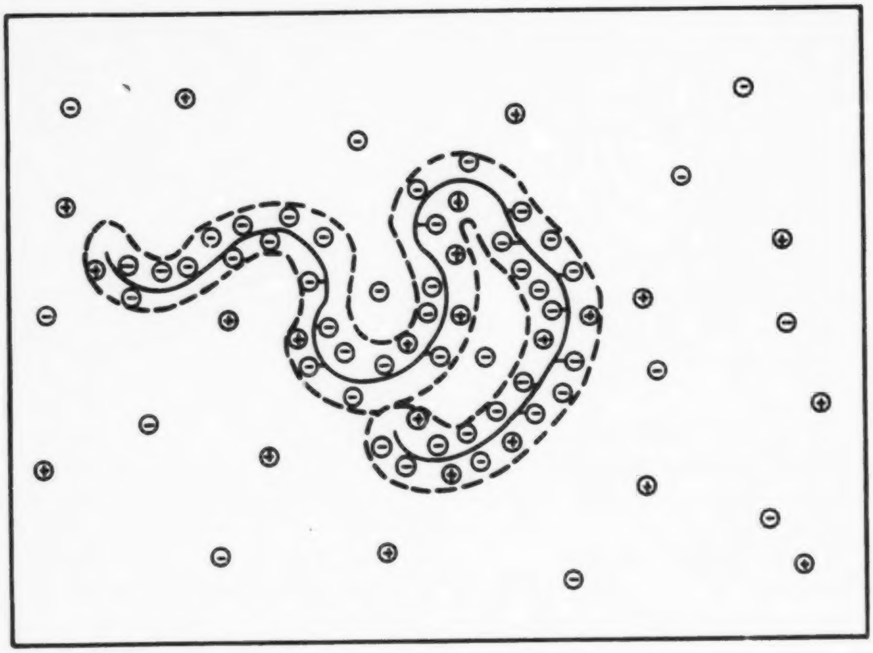


FIGURE 6-5. Typical configuration of a polyelectrolyte in solution [3].

Since an extensive variety of widely different types of polymers are currently in use, it is impossible to generalize about the applicability of polymers to particular sludges. Laboratory or pilot scale evaluations are essentially always required. The WPCF *Manual of Practice on Sludge Dewatering* [4] has very aptly noted that seldom do any two polymers give the same results, some polymers work best in conjunction with other chemicals, some polymers may coagulate well but still not improve sludge dewaterability. Further, the polymer that best improves a sludge's settleability may not be the same one that improves its dewaterability.

6.3.2 Use of Inorganic Chemicals

Polyvalent metal ions (ferric, ferrous, and aluminum) hydrolyze in water to produce polynuclear complexes. The important metal salts used in sludge conditioning are ferric chloride and ferrous sulfate. They function primarily as coagulants. The hydrolyzed salts possess a significant charge and some polymeric properties as well. Accordingly, they provide charge neutralization and enmeshing capabilities toward dispersed material.

Hydrated lime is almost always used in conjunction with metal salts. Though lime does have some slight dehydration effect on colloids, its use in conditioning is also for pH control, odor reduction, disinfection, and filter aid effect.

Both power plant fly ash and sludge incinerator ash have been used successfully in the conditioning of sludge. The properties of ash which enable it to improve dewatering of sludge include partial solubilization of its metallic constituents, its sorptive capabilities, and its irregular particle sizes [5]. The city of Indianapolis, Indiana, recently started using sludge incinerator ash to condition a mixture of primary and activated sludges prior to dewatering. The dramatic effect of ash addition on the average performance of the plant's rotary vacuum belt filters is shown in Table 6-6. Indianapolis has succeeded in increasing its filter productivity by as much as 500 percent and decreasing its cake moisture by as much as 22 percent. The filtrate quality has also been improved and from a cost viewpoint the cationic polymer requirement has been reduced by approximately 55 percent. These data were obtained in late 1972 and at that time the plant was handling approximately 418,000 lb of dry sludge per day and the ash to dry sludge solids ratio varied from 0.25 to 0.50. The ash handling facilities required almost zero investment and no additional operating cost. The relative location of the ash slurry line to the gravity sludge thickeners provided for easy ash addition requiring only the installation of a tap, short feed line, and a pump [6].

6.3.3 Elutriation

Elutriation is essentially a washing process once widely used for conditioning anaerobically digested sludges prior to further conditioning with a metal salt. The process involves countercurrent or cocurrent extraction of the soluble alkaline carbonates and phosphates as well as fine sludge particles from the sludge by dilution with treatment

TABLE 6-6

EFFECT OF ASH ADDITION ON VACUUM FILTRATION AT INDIANAPOLIS

Parameters	Before Ash Addition	After Ash Addition
Filter Yield (lb D.S./ft^2/hr)	1.0 – 2.0	5.52 – 13.0
Cake Moisture (%)	85	66.5
Filtrate Quality	Poor	Excellent
Filter Capacity	Insufficient	More than Adequate
Cationic Polymer Required (lb/ton)	15	6.8
Release of Cake from Media	Poor	Excellent
Supplemental Fuel Requirement	Little or None	Little or None

plant effluent and resettling. The principal purposes of the process are to reduce chemical requirements and produce a more readily dewaterable sludge. With the advent of higher levels of secondary treatment and consequent activated sludges, the sludge going to elutriation contains a large amount of fine particles. Therefore, the process, unless flocculants are used, will produce a very dirty elutriate and a heavy recirculation load. Plant studies will be reviewed in Chapter 10 which indicate a need for a postdigestive thickening process like elutriation in plants processing combined primary and secondary sludges.

6.3.4 Heat Treatment

In heat treatment, temperatures of from 300 to 500° F and pressures of 150 to 400 psig are attained for protracted periods. Significant changes in the nature and composition of wastewater sludges result. The effect of heat treatment has been ideally likened to syneresis, or the breakdown of a gel into water and residual solids. Wastewater sludges are essentially cellular material. These cells contain intracellular gel and extracellular zoogleal slime with equal amounts of carbohydrate and protein. Heat treatment breaks open the cells and releases mainly proteinaceous protoplasm. It also breaks down the protein and zoogleal slime, producing a dark brown liquor consisting of soluble polypeptides, ammonia nitrogen, volatile acids, and carbohydrates. The solid material left behind is mineral matter and cell wall debris.

Dewaterability is improved by the solubilizing and hydrolyzing of the smaller and more highly hydrated sludge particles which then end up in the cooking liquor. While analysis of this liquor from domestic wastewater sludges indicates the breakdown products are mostly organic acids, sugars, polysaccharides, amino acids, ammonia, etc., the exact composition of the liquor is not well defined [7]. Corrie and Wycombe [8] have found the liquor to be highly polluted and contain a high proportion of nonbiodegradable matter. This matter is largely humic acids, which can give rise to unpleasant odors and taste if present in water which has been chlorinated prior to use for domestic supply. If industrial wastes of various types are included in the wastewater to be treated, the actual chemical composition of the liquor resulting from heat treatment of the sludge should be determined by a detailed chemical analysis. The Thames Conservancy District of England requires when heat treatment of sludge is employed in a plant that the liquor be treated separately. The treatment process chosen must include adsorption on activated carbon of the COD. The reasons for these requirements are that too little is as yet known about the exact composition of soluble organic matter in the liquor to permit its recycle into the Thames River since it is used as a downstream water supply.

Figure 6-6 gives the conceptual design recommended by the District for handling heat treatment liquor based on their pilot plant efforts.

A review of reported analyses [8,9] of liquor from the heat treatment of sludge gives the range of values shown:

BOD_5 = 5,000 to 15,000 mg/l

COD = 10,000 to 30,000 mg/l

Ammonia = 500 to 700 mg/l

Phosphorus as P = 150 to 200 mg/l

About 20 to 30 percent of the COD is not biodegradable in a 30-day period. The volume of cooking liquor from an activated sludge plant with heat treatment amounts to 0.75 to 1.0 percent of the wastewater flow. Based on BOD_5 and solids loadings, the liquor can represent 30 to 50 percent of the loading to the aeration system. The pH of cooking liquors is normally in the range of 4 to 5, which necessitates chemical neutralization and/or corrosion resistant equipment. Work by Erickson and Knopp [10] presents data indicating few problems with treatment of cooking liquors from Zimpro systems.

6.4 Physical Factors in Conditioning Processes

Plant experiences have shown that the conditioning requirements and hence the performance achieved in thickening and dewatering processes are affected by the manner in which sludge is treated.

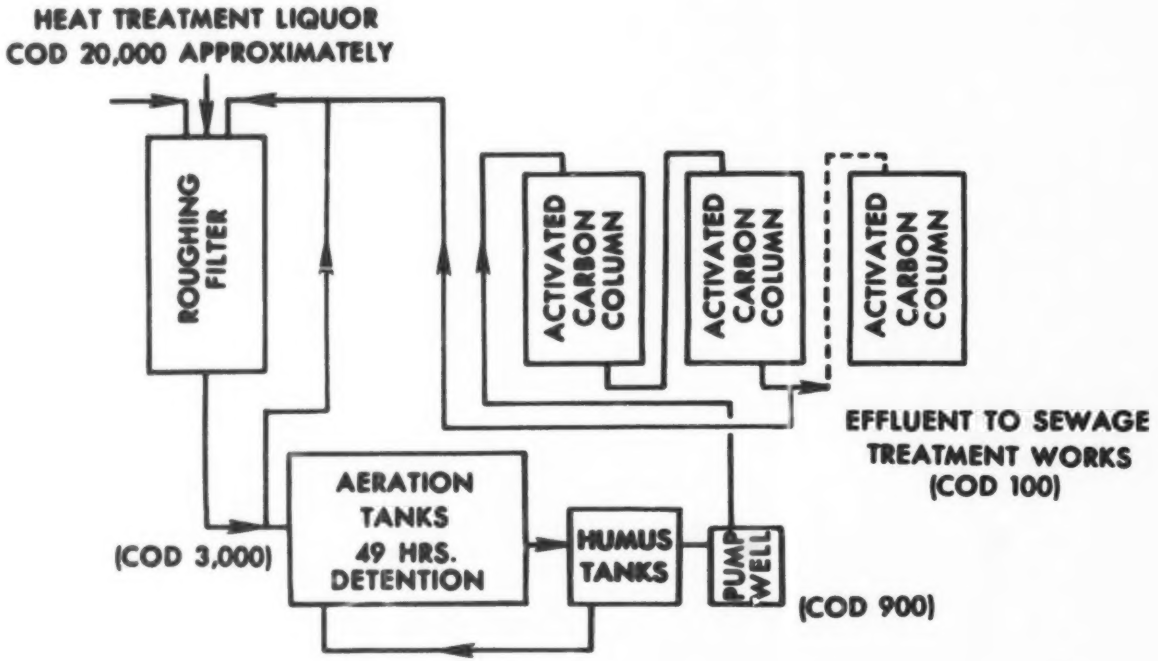

HEAT TREATMENT LIQUOR
COD 20,000 APPROXIMATELY
ROUGHING FILTER
ACTIVATED CARBON COLUMN
ACTIVATED CARBON COLUMN
ACTIVATED CARBON COLUMN
EFFLUENT TO SEWAGE TREATMENT WORKS (COD 100)
AERATION TANKS 49 HRS. DETENTION
(COD 3,000)
HUMUS TANKS
PUMP WELL
(COD 900)

FIGURE 6-6. Schematic diagram of plant for processing heat treatment liquor.

6.4.1 Effect of Processing Prior to Conditioning

Any operation which tends to irreversibly disaggregate, hydrate, or subject sludge to high shear conditions is deleterious to solid-liquid separation procedures. One striking example of this tendency is the effect noted when pumping sludges through pipelines from satellite plants to a central dewatering location. It has been demonstrated that sludge is easily and economically dewaterable prior to piping, while the sludge exiting from the pipeline is not easily dewaterable.

Other treatment plant operations, including aeration and thickening, also subject the sludge particles to shear forces. Although these operations normally have minimal shear effects, there is some adverse speculation about the use of surface aerators and their effect on floc characteristics. Unfortunately the results of no large scale comparisons between surface and diffused aeration and their effects on floc processability have as yet been reported.

6.4.2 Conditioner Application

The optimum sequence of addition is best determined by trial and error when two or more conditioners are used. Normally ferric chloride is added first when used with lime. Anionic polymer is added first when a combination of anionic and cationic polymer is needed. In cases where experimentation is not possible, it is best to provide flexibility in design of the system. In practically all cases, the cardinal rule for design of equipment for mixing conditioning chemicals with sludge is to provide just enough mixing to disperse the conditioner throughout the sludge. This is necessary to minimize floc shearing. Visual inspection capability should be provided for the sludge-flocculant mixture from the point of their contact on into the dewatering or thickening unit. A further design consideration is the provision of individual conditioning units for each dewatering unit. It is not always economical to provide one common conditioning unit for several dewatering units. Problems can arise in balancing the flow rates of the various streams when starting up or shutting down individual units and in locating the conditioning unit relative to each dewatering device.

One of the most widely used types of conditioning units is the rotating horizontal drum with internal baffles, which resemble a cement mixer. Another is the baffled open trough with paddle mixers. Vertical cylindrical tanks with propeller mixers were used at one time, but these have not been popular in recent years. The rotating horizontal drum type of conditioner is depicted in Figure 6-7. While this type unit normally functions well, care should be exercised in minimizing the vertical height differential between the point of discharge from the drum and the vat level. This is particularly important with thin, shear-sensitive sludges. It has been helpful in some cases to baffle the point of discharge from the chute into the vat for the prevention of localized wash-off of filter cake.

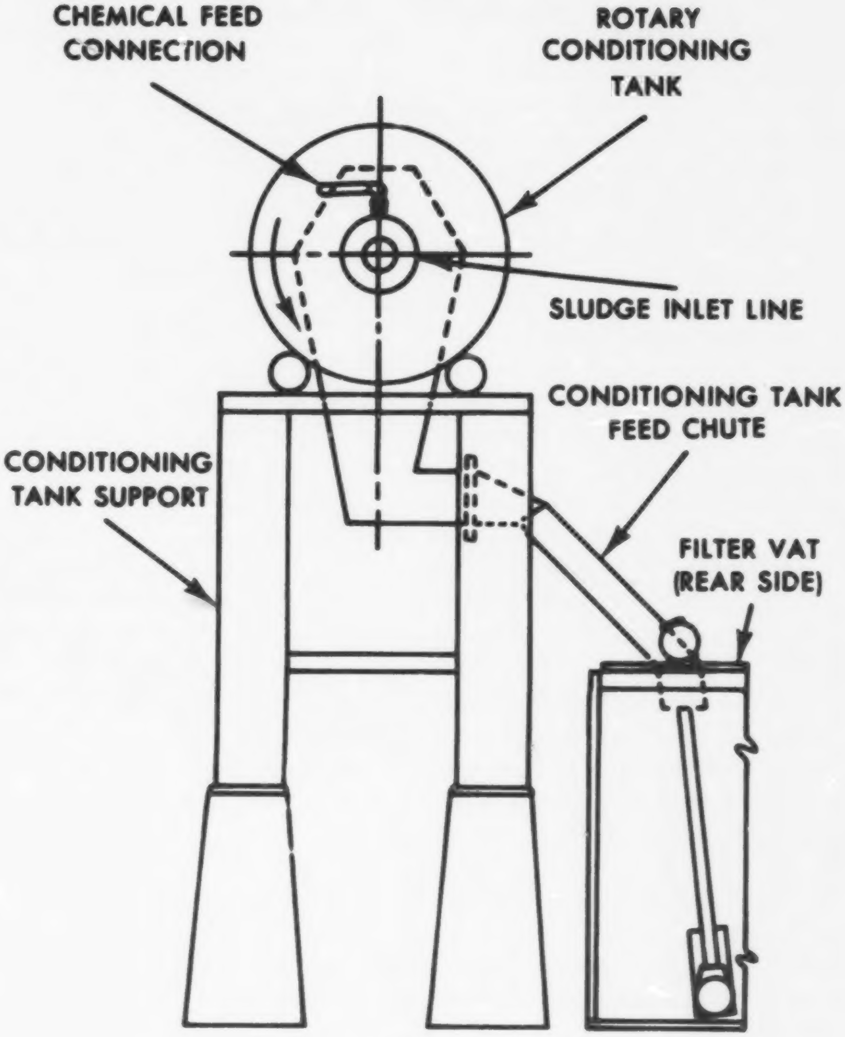

CHEMICAL FEED CONNECTION
ROTARY CONDITIONING TANK
SLUDGE INLET LINE
CONDITIONING TANK FEED CHUTE
CONDITIONING TANK SUPPORT
FILTER VAT (REAR SIDE)

FIGURE 6-7. Rotary drum conditioner.

Courtesy Komline-Sanderson

A sectional view of the baffled trough type conditioning unit appears in Figure 6-8. This type unit has two important features. It permits continual visual inspection of the conditioning operation and the shear imparted to the conditioned sludge as it flows into the vat can be minimized.

6.5 Conditioning for Gravity Thickening

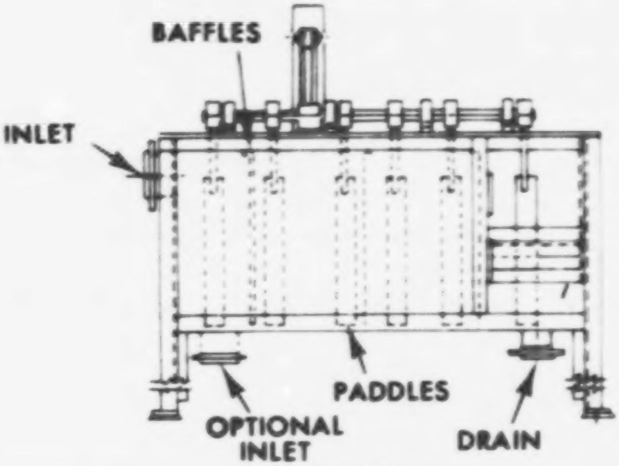
Primary sludges and mixtures of primary and trickling filter sludges normally respond to gravity thickening without conditioning. However, mixtures of primary sludge and EAS can present a problem for gravity thickening. Flocculants are required to ensure good solids capture and loading rate, and even then as was discussed in Chapter 4, a high underflow solids concentration is difficult to obtain. It is generally preferable to separately DAF thicken excess activated sludge. When flocculants are used to condition sludge for gravity thickening, the flocculant solution should be added to either the sludge or the dilution water on its way into the thickener. Recent experience indicates that excess oxygen activated sludge may be amenable to gravity thickening [11]. Data on an actual plant operation is presented in Chapter 10.

6.6 Conditioning for Flotation Thickening

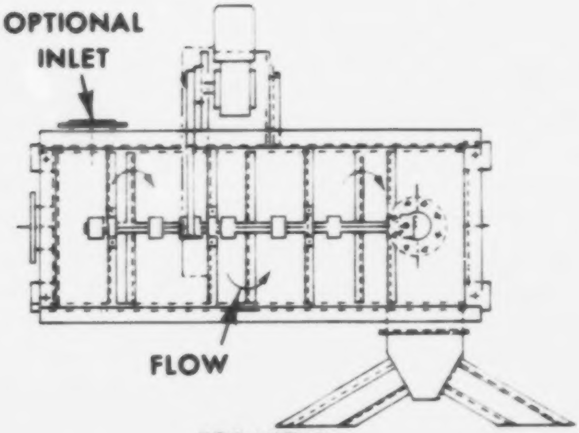
Chemicals can assist flotation by increasing the solids loading rate, float cake solids concentration, and solids capture. The first two parameters are interrelated, and this dependency as well as the effect of polymer addition is illustrated by Figure 6-9 for an activated sludge.

Since a minimum practical cake solids concentration in the float is usually 4 percent, the operable conditions for this particular activated sludge and system are in the range of 4 to 6 lb of polymer per ton of dry solids and mass loadings of 10 to 30 lb/day/sq ft. While such loading rates compare favorably with those of a gravity thickener, flotation units normally achieve loading rates of 48 to 96 lb/day/sq ft with 1 to 5 lb/ton of a chemical and resultant cake solids concentrations of 4 to 6 percent solids.

Figure 6-10 shows a typical layout for addition of chemical flotation aids. The chemical is usually added at the mixing chamber where the pressurized recycle flow is mixed with the sludge stream. The design of the inlet mixing chamber and point of polymer application are very important. Injection of the flotation aid solution into the recycle line just as the bubbles are being formed and mixed with the sludge produces the best results [12]. This assures excellent mixing, minimum sludge particle shear, and positive air bubble adsorption.


BAFFLES
INLET
PADDLES
OPTIONAL INLET
DRAIN

SIDE VIEW


OPTIONAL INLET
FLOW

TOP VIEW

FIGURE 6-8. Baffled trough unit.

Courtesy Rexnord

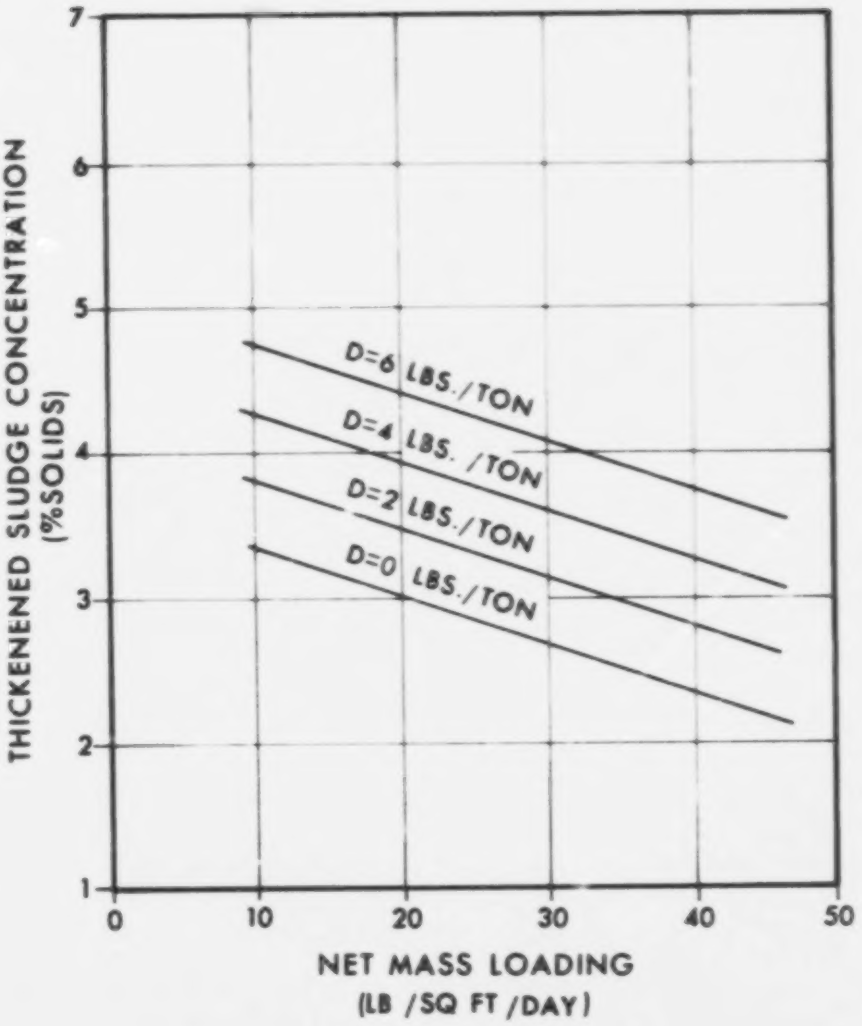


FIGURE 6-9. Thickening performance as affected by mass loading at constant chemical dosage. *Courtesy Rexnord*

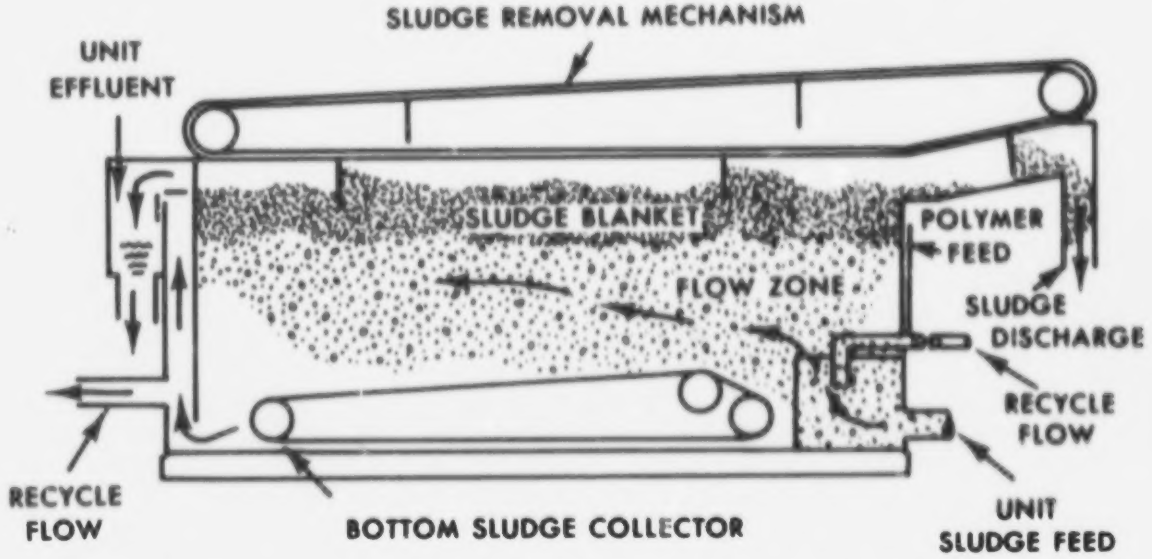

SLUDGE REMOVAL MECHANISM
UNIT EFFLUENT
SLUDGE BLANKET
POLYMER FEED
FLOW ZONE
SLUDGE DISCHARGE
RECYCLE FLOW
RECYCLE FLOW
BOTTOM SLUDGE COLLECTOR
UNIT SLUDGE FEED

FIGURE 6-10. Flow diagram of a flotation unit.

Courtesy Komline-Sanderson

6.7 Conditioning for Dewatering

An important operating variable in dewatering is the conditioning chemical and dosage. Optimum performance of the dewatering equipment hinges on a determination of the most economical and effective conditioning method. Conditioning for improved dewatering by vacuum filtration, centrifugation, drying beds, and pressure filtration is discussed.

6.7.1 Rotary Vacuum Filtration

Items of primary concern in conditioning sludge for feed to a rotary vacuum filter include:

- Cake pickup on the drum
- Production rate
- Cake solids content
- Cake discharge from filter media
- Filtrate quality (solids capture)

The determination of optimum conditioning procedures hinges on trial and error plant testing. However, plant scale testing is often preceded by laboratory and pilot scale testing to narrow down the selection of conditioning systems and levels of operation. The two most widely employed laboratory tests are the Buchner funnel and filter leaf [4,13]. Pilot plant testing is generally conducted with a small model (3 ft diameter with a 1 ft wide face) of a plant scale rotary vacuum filter. Most manufacturers will rent pilot equipment for test purposes. Extended testing of a limited number of operating modes can then be conducted on a plant scale.

Table 6-7 shows estimated chemical requirements for optimum vacuum filtration of a variety of sludges. The estimates are based on experiences at different treatment plants throughout the United States and are listed here only for guidance [14]. All of the various types of conditioning procedures are used in conjunction with rotary vacuum filters.

6.7.2 Centrifuges

Polymeric flocculants are normally used for sludge conditioning in solid bowl centrifuges, which are the prevalent type used. The metal salts are generally not used because of corrosion problems. Principal performance parameters for gauging effectiveness of conditioners are production rate, cake solids content, and centrate solids content. A previous deterrent to widespread use of continuous horizontal solid bowl centrifuges was

TABLE 6-7

ESTIMATED CHEMICAL CONDITIONING DOSAGE FOR VACUUM FILTRATION

Type of Sludge	CaO Dose (lb/ton)	$FeCl_3$ Dose (lb/ton)	CaO + $FeCl_3$ ($/ton)	Polymer Dose (lb/ton)	Polymer ($/ton)
Primary Sludge	176	42	5.01	5	1.65
Limed Primary (212 lb CaO/ton)	0	42	2.81	5	1.65
Digested Primary Sludge	240	76	7.86	20	6.60
Digested/Elutriated Primary	0	68	4.56	9	2.97
Raw (Primary + EAS)	200	52	5.98	18	5.94
Limed (Primary + EAS)	0	40	2.68	5	1.65
Digested (Primary + EAS)	372	110	12.02	36	11.88
Digested/Elutriated (Primary + EAS)	0	125	8.38	24	7.92

Note: CaO cost = $0.0125/lb

$FeCl_3$ cost = $0.067/lb

Polymer cost = $0.33/lb

the problem of providing adequate solids capture with reasonable production rates and flocculant dosages. The crux of that problem was attainment and maintenance of floc stability in the face of high shear conditions. This problem has been minimized by direct injection of the flocculant into the centrifuge to avoid exposing flocculated material to high shear. This design change reduced flocculant consumption and made the process more efficient and competitive. Figure 6-11 shows a cross section of a typical centrifuge with the polymer injection arrangement. The point labelled floc nozzle is the application point for the polymer flocculant. Recently proper flocculation has permitted the centrifuge to perform well at low rotational speeds. This reduced conditioning chemical requirements and maintenance due to wear. Optimum conditioning can best be determined in pilot tests.

6.7.3 Drying Beds

Use of conditioning procedures with sludge drying beds is not widespread; however, elutriation and polymers are employed in isolated cases. Mogelnicki [14] details experiences with elutriation and polymer conditioning.

Conditioning agents are used with sludge drying beds at the Chicago Southwest Plant to reduce drying time and maximize bed production during fair weather. Approximately 0.5 lb of cationic polymer per ton of dry solids is effective in decreasing the drying time from 13 to 5 days. Reported cost of the polymer is about $0.50/dry ton of sludge.

Figure 6-12 presents data from a series of tests with a digested sludge having a solids concentration of 4 percent. This sludge was conditioned with a cationic polyelectrolyte. The effect of various levels of polymer addition on drainage time and the time elapsed before the cake can be readily lifted from the bed can be easily seen. The use of sludge conditioning in this case produced significant increases in both drainage rate and ultimate cake solids content [14].

6.7.4 Filter Presses

Filter aids such as ash and inorganic conditioners are used in dewatering operations with filter presses. Presses depend on the exertion of massive pressures ($\leq$ 200 psi) to squeeze water out of sludge. Consequently, conditioning problems are more difficult than with other methods and laboratory and/or pilot plant evaluation is needed. These high pressures tend to destroy the flocculation achieved with normal conditioning. Accordingly, relatively large doses of lime or recycled ash (1.0 to 1.5 parts ash/part dry solids), with or without metal salts, are used. The Sheffield, United Kingdom [15] plant uses 27.5 percent lime and 13 percent Fe_2O_3 on a sludge solids basis.

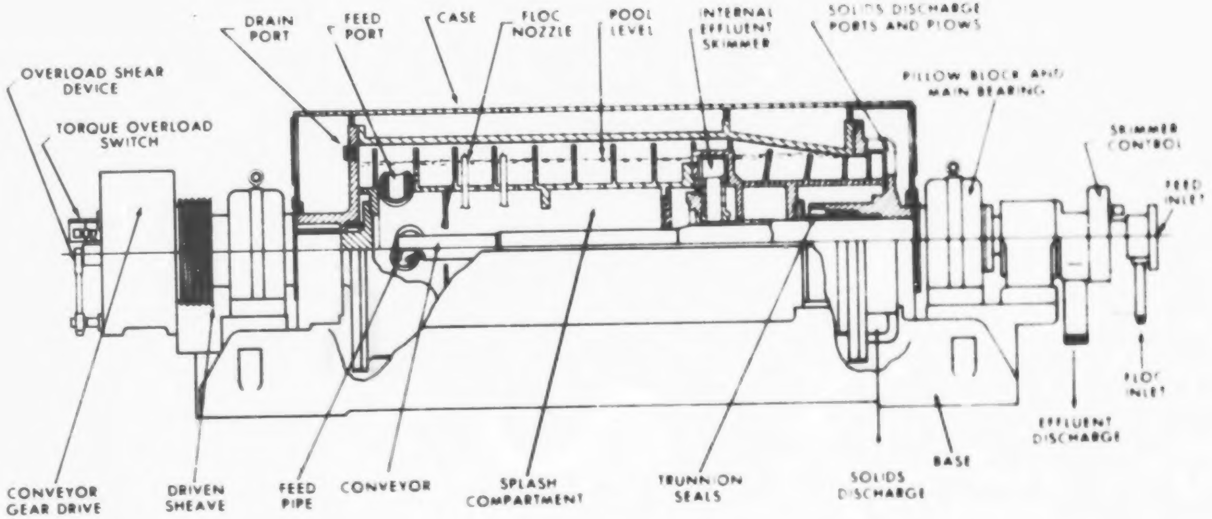

OVERLOAD SHEAR DEVICE
TORQUE OVERLOAD SWITCH
DRAIN PORT
FEED PORT
CASE
FLOC NOZZLE
POOL LEVEL
INTERNAL EFFLUENT SKIMMER
SOLIDS DISCHARGE PORTS AND PLOWS
PILLOW BLOCK AND MAIN BEARING
SKIMMER CONTROL
FEED INLET
FLOC INLET
EFFLUENT DISCHARGE
BASE
SOLIDS DISCHARGE
TRUNNION SEALS
SPLASH COMPARTMENT
CONVEYOR
FEED PIPE
DRIVEN SHEAVE
CONVEYOR GEAR DRIVE

FIGURE 6-11. Concurrent flow solid-bowl centrifuge

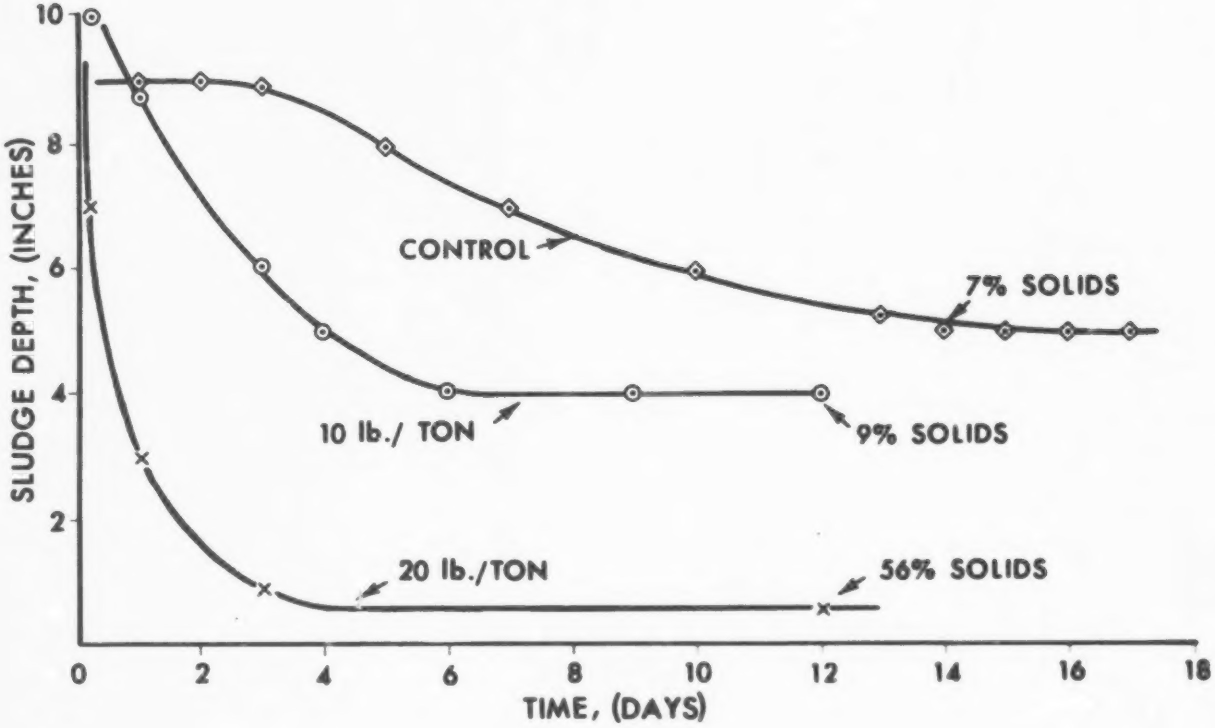


FIGURE 6-12. Sandbed dewatering [14]

6.8 Selection of Conditioning Chemicals

It is important for the design engineer, in evaluating alternate types of conditioning, to become more familiar with the available chemicals. Table 6-8 lists several available chemicals and indicates a broad price range.

TABLE 6-8

CONDITIONING CHEMICAL MATERIALS

Type	Form	Price (\$/lb)*	Suppliers
$FeCl_3$	Liquid	0.06	Two major, Several minor
Cationic Polymer	Dry Powder	0.50-1.50	About 10
Cationic Polymer	Liquid	0.05-0.50	About 10
Anionic Polymer	Dry Powder	0.60-1.30	About 15

*Based on 1973 costs figures.

The significant points to consider in selecting these conditioning chemicals are:

- Ferric chloride is a commodity chemical, made either from waste pickle liquor or more usually from scrap metal and chlorine. With proper handling procedures and equipment, it is not difficult to employ in the normal 40 percent solution form. Since it is a low priced commodity chemical, its analysis and quality can be easily controlled with relatively simple analytical control procedures.
- The price structure of ferric chloride is relatively stable.
- Cationic polymeric flocculants, which are specialty chemicals, are available in a wide variety. The chemical composition, functional effectiveness, and cost-effectiveness of the various products differ greatly. Accordingly, the products of several different companies should be evaluated to optimize efficiency.
- Because of the wide variety and divergent prices of the various polymers, one should always compare them in terms of cost per ton of sludge solids conditioned rather than in pounds of polymer required per ton.

- Since polymers are specialty chemicals, their production and composition are the subject of continuing research and development by the suppliers. This has usually resulted in continuing improvements in functional effectiveness and cost.

6.9 References

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CHAPTER 7

SLUDGE DEWATERING

7.1 Methods and Functions

The methods used to remove sufficient water from liquid sludges so as to change the physical form to that of a damp solid are best described in terms of the particular type of dewatering device used. The commonly used devices include:

- Rotary vacuum filters
- Centrifuges
- Drying beds
- Lagoons
- Filter presses
- Horizontal belt filters

The relationship of the various dewatering methods to those processes which immediately precede and follow them are summarized in Table 7-1.

An ideal dewatering operation would capture practically all the solids in the dewatered cake at minimum cost. The resultant cake would have the physical handling characteristics and moisture content optimal for subsequent processing. Process reliability, ease of operation, and compatibility with the plant environment would also be optimum.

The technology and design of all available dewatering methods is constantly under development, particularly in the past five years. Each type, therefore, should be given careful consideration. The applicability of a given method should be determined on a case-by-case basis with the specifics of any given situation being carefully evaluated, preferably in pilot tests.

TABLE 7-1

THE RELATIONSHIP OF DEWATERING TO OTHER SLUDGE TREATMENT PROCESSES FOR TYPICAL MUNICIPAL SLUDGES

Method	Pretreatment Normally Provided		Normal Use of Dewatered Cake			
	Thickening	Conditioning	Landfill	Land Spread	Heat Drying	Incineration
Rotary Vacuum Filter	Yes	Yes	Yes	Yes	Yes	Yes
Centrifuge (Solid Bowl)	Yes	Yes	Yes	Yes	Yes	Yes
Centrifuge (Basket)	Variable	Variable	No	Yes	No	No
Drying Beds	Variable	Not Usually	Yes	Yes	No	No
Lagoons	No	No	Yes	Yes	No	No
Filter Presses	Yes	Yes	Yes	Variable	Not Usually	Yes
Horizontal Belt Filters	Yes	Yes	Yes	Yes	Yes	Yes

7.2 Rotary Vacuum Filtration

7.2.1 Mechanics of Rotary Vacuum Filtration

Comprehension of the theoretical aspects of rotary vacuum filtration of wastewater sludges plus practical application of the theory through the medium of lab, pilot, and full-scale plant test procedures is essential in evaluating systems. The clearest and most up-to-date expositions on the mechanisms of rotary vacuum filter dewatering have been made by Bennett and Rein [1], Bennett, Rein, and Linstedt [2], and Gale [3,4,5].

Theory

Figure 7-1 shows a sectional view of a rotary filter which consists of a cylindrical drum rotating partially submerged in a vat or pan of conditioned sludge. The drum is divided radially into a number of sections, which are connected through internal piping to ports in the valve body (plate) at the hub. This plate rotates in contact with a fixed valve plate with similar parts, which are connected to a vacuum supply, a compressed air supply, and

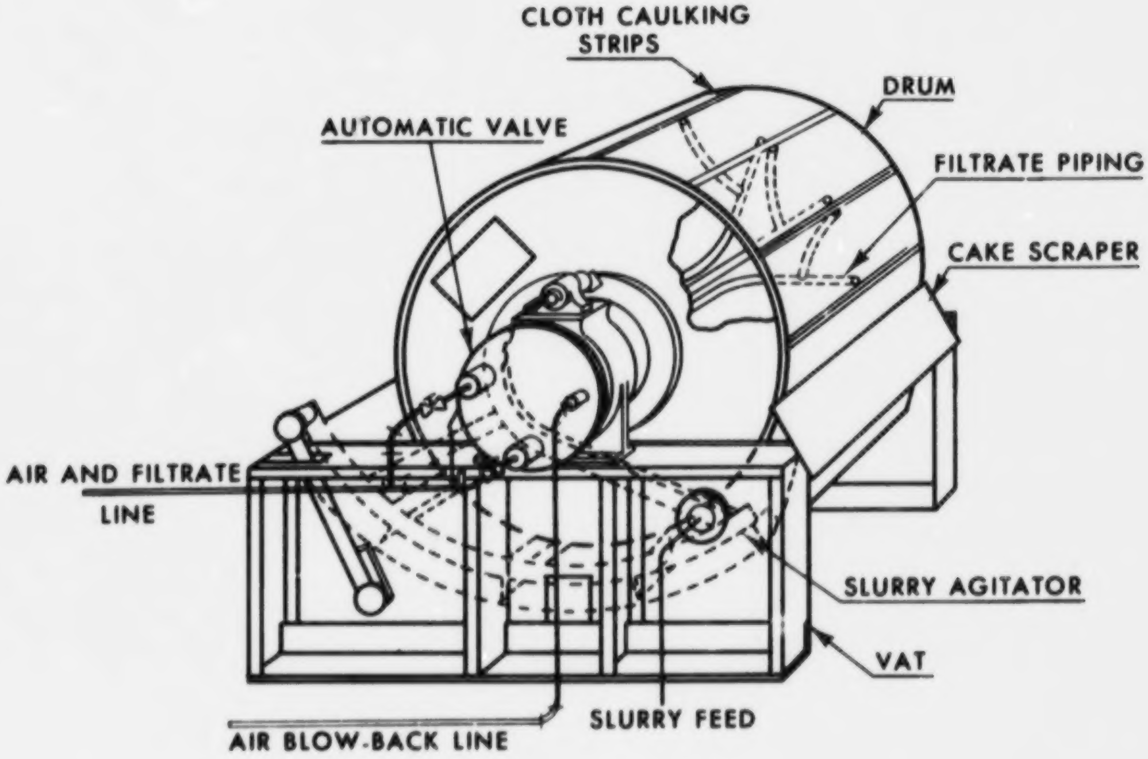
CLOTH CAULKING STRIPS
DRUM
AUTOMATIC VALVE
FILTRATE PIPING
CAKE SCRAPER
AIR AND FILTRATE LINE
SLURRY AGITATOR
VAT
SLURRY FEED
AIR BLOW-BACK LINE

FIGURE 7-1. Cutaway view of a rotary drum vacuum filter.

an atmospheric vent. As the drum rotates each section is thus connected to the appropriate service. Figure 7-2 illustrates the various operating zones encountered during a complete revolution of the drum. In the pickup or form section, vacuum is applied to draw liquid through the filter covering (media) and form a cake of partially dewatered sludge. As the drum rotates the cake emerges from the liquid sludge pool, while suction is still maintained to promote further dewatering. A lower level of vacuum often exists in the cake drying zone.

Continuous filtration is a cyclic process and operation encompasses various rate functions. Furthermore, while one of the rate functions normally may be controlling, all interact and, therefore, none can be ignored. Two distinct rate phenomena are encountered in continuous operation of vacuum filters for sludge dewatering and warrant special attention [6]. They are filter sludge cake formation rate and dewatering of filter cake to obtain the desired final moisture content.

The hydraulics of filtrate flow were developed by Ruth, Motillon, and Montonna [7] and Carmen [8] using Darcy's law and the Carmen-Kozeny equation. This approach was adapted to wastewater sludge filtration by Coakley and Jones [9], Jones [10], and Hatscheck [11], Halff [12], and Grace [13]. The theory is based on several assumptions. These include the laminar flow condition, a constant volume of solids deposition with each increment of filtrate, and a constant increase in filtrate flow resistance for each volume of cake solids deposited. With these conditions, an average specific resistance for each unit thickness of cake deposited can be assumed. Using the average specific resistance concept, the following equation has been developed [6]. This average specific resistance (r) is the resistance of a unit weight of cake per unit area at a given pressure and is expressed in $\sec^2/g$.

$$r = \frac{2PA^2b}{\mu w}$$

where

r = Average specific cake resistance (usually constant for one slurry), $\sec^2/g$

P = Pressure drop through filter medium and sludge cake, cm of water

A = Area of filtering surface, cm^2

b = Slope of the t/V vs. V plot in sec/ml^2

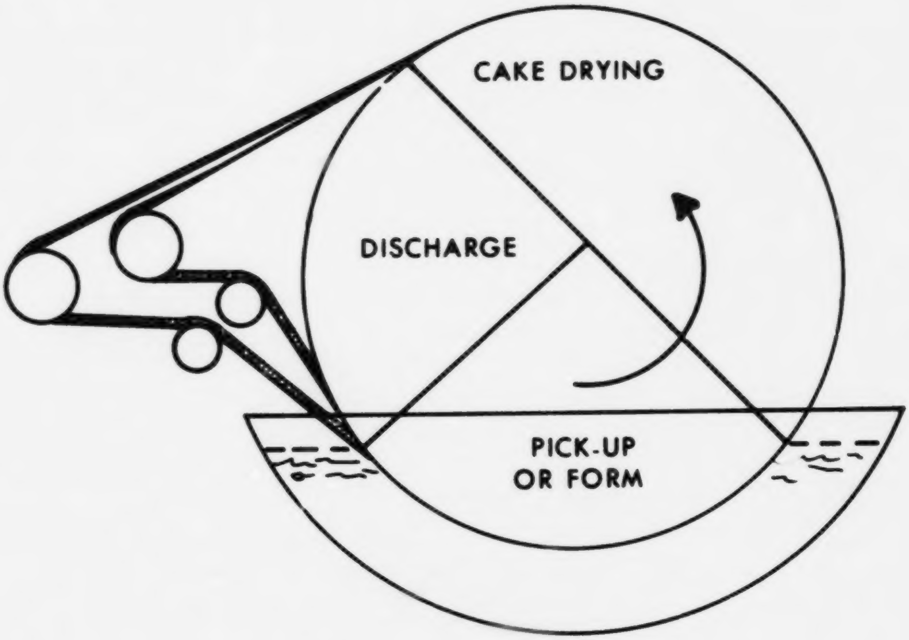
CAKE DRYING
DISCHARGE
PICK-UP
OR FORM

FIGURE 7-2. Operating zones of a vacuum filter.

μ = Viscosity of filtrate in poise

w = Weight of dry sludge cake solids per unit volume of filtrate, gm/cc

A value is obtained for "b" by conducting a simple Buchner funnel test and this is discussed in a later section. Some derivations include the resistance of the filter medium in series with the cake resistance. Normally, the filter media resistance is negligible, and the term can be dropped [6]. The average specific resistance (r) is assumed to be constant for any one slurry and operating condition. It can be altered by the application of conditioning techniques and it is a function of the vacuum level applied. Most municipal sludge solids deform at high vacuum levels and fill the pore openings, which increases the resistance per unit of vacuum. This can be expressed as:

$$r = r'P^s$$

where

r' = Constant representing the specific resistance of a noncompressible cake

s = Cake compressibility exponent

The cake compressibility coefficient (s) varies from zero for a rigid incompressible cake to greater than one for highly compressible cakes. For domestic wastewater sludges, this value ranges from 0.4 to 0.85 [6].

The following equation relates the specific resistance to the yield of a rotary vacuum filter.

$$Y = \left(\frac{2P}{\mu} \cdot \frac{W}{\bar{r}_p} \cdot \frac{F_f}{\theta_R}\right)^{1/2} F_c$$

where

Y = Yield of filter in mass of dry suspended solids produced in unit time per unit of total area of filter medium

P = Pressure difference across filter cake during cake formation

W = Mass of dry suspended solids per unit volume of liquid in sludge

F_f = Fraction of total filter area used for cake formation

μ = Viscosity of filtrate

$\bar{r}_p$ = Apparent specific resistance of cake measured at P

θ_R = Time for one revolution of the filter

F_c = Cake correction factor; the ratio of the mass of liquid in unit mass of sludge, to the mass of filtrate obtained when unit mass of sludge is filtered

This equation allows prediction of expected data with changes in pressure, feed concentration, viscosity, and the cake formation time [3]. For example, if the length of the cake formation time is quadrupled, the filtration rate is cut in half. An increase in the temperature of the slurry, in general, decreases the viscosity and increases the cake formation rate. As the filter feed solids concentration is increased, the solids rate also increases. As the feed solids concentration increases, many filter cakes exhibit a more permeable bridging and a subsequent reduction in the cake resistance. It has been noted that the influence of feed solids concentration on the r value cannot be predicted from theory, and thus parameters of feed solids concentration also must be employed in the correlation methods. Physically, another factor comes into consideration, that is, because a filter is a machine using energy to separate solids from liquids, the less liquid there is to remove, the higher will be the rate of dry solids production. Empirically it has been shown that the dry solids production rate is for all practical purposes directly proportional to the feed solids concentration. Data for several experiments at a cycle time of 120 seconds are plotted in Figure 7-3 on logarithmic scales. In this figure, the measured yields corrected for sludge solids content are plotted against specific resistance. The data are seen to closely lie on a straight line of slope -0.5, as theory would predict.

Test Procedures for Sizing Vacuum Filters

The two test procedures used for determining the filterability of sludges are the Buchner funnel method and the filter leaf technique. The Buchner funnel method enables a determination of the relative effects of various chemical conditioners and the calculation of the specific resistance of the sludge, but it is seldom used for the calculation of required filter area. The filter leaf test is used to determine the required filter area.

One laboratory system using a machined aluminum Buchner funnel apparatus is shown in Figure 7-4. Typically, two pieces of No. 4 Whatman filter paper are fixed in the Buchner

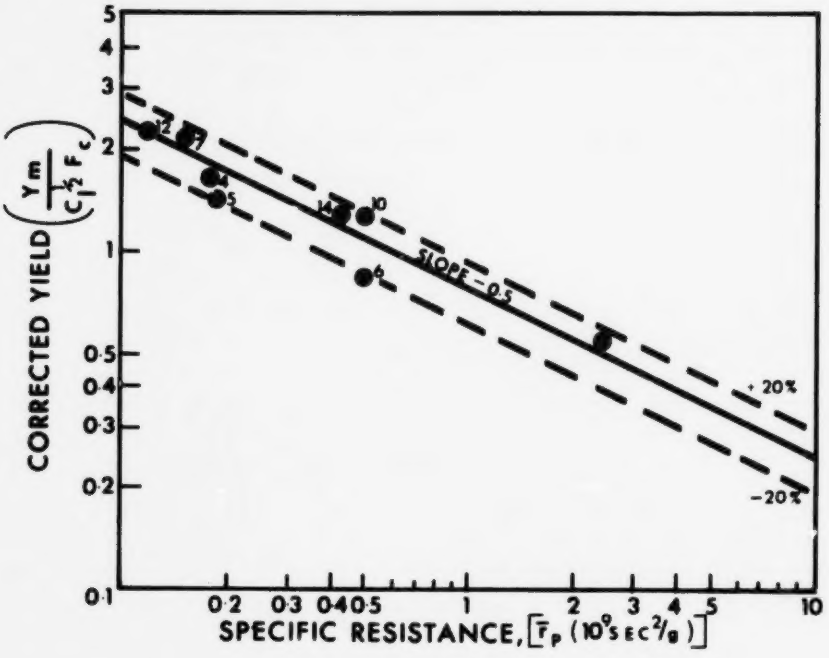

FIGURE 7-3. Corrected filter yield vs. specific resistance [4].

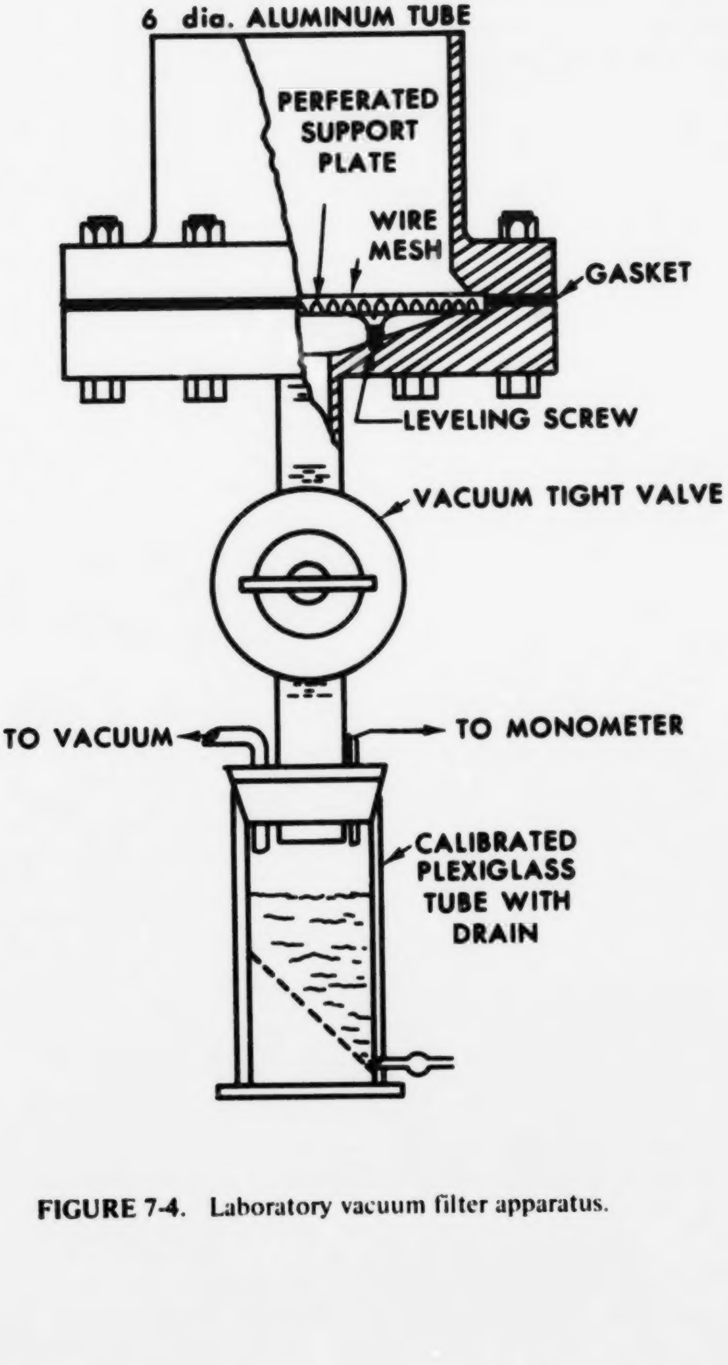
6 dia. ALUMINUM TUBE
PERFERATED SUPPORT PLATE
WIRE MESH
GASKET
LEVELING SCREW
VACUUM TIGHT VALVE
TO VACUUM
TO MONOMETER
CALIBRATED PLEXIGLASS TUBE WITH DRAIN

FIGURE 7-4. Laboratory vacuum filter apparatus.

funnel and the sludge to be tested is introduced in a single large batch. The depth of sludge used is about 1½ cm which is approximately equal to the amount of sludge filtered in a single pass of a rotating filter. The valve on the vacuum line is opened to initiate filtration. The accumulative volume of filtrate is recorded at appropriate time intervals. A plot of t/V as a function of V (as shown in Figure 7-5) permits the calculation of the specific resistance. Typical values of specific resistance vary from 5×10^7 to 70×10^7 sec^2/g for conditioned sludges.

A drawback of this type of testing is that the time and volume are taken on an accumulative basis, which tends to underemphasize certain important portions of the curve. Crook and Jones [14] have shown that the upper end of the curve includes a plateau prior to the drying phase. This can be observed when the test is carried out in such a way that the instantaneous flow rate, dV/dt or $1/Q_i$, its inverse, is measured instead of the accumulated average flow rate. A plot of the inverse of instantaneous flow rate as a function of the volume of sludge applied is shown in Figure 7-6. When instantaneous flow rates are used, the initial portion of the curve is also affected, in the manner shown. Figure 7-7 shows the correlation between the filtrate flow plot and the normal operating cycle of a rotating vacuum filter. Four different processing phases exist. In Phase I, solids capture increases from near zero percent, just as the media contacts the sludge slurry, to near 100 percent capture at the end of the phase. Phase II is the continuation of cake formation under the conditions of nearly complete solids capture. Phase III occurs immediately after the sludge coated media leaves the vat and is characterized by water exiting from the larger capillary pore openings. In Phase IV the cake is further dewatered by air drying. Phase II is the only portion of the curve that follows theory.

The Buchner funnel test enables a prediction of the effects of various conditioning chemicals. It does not permit a precise estimate of filter size and operating characteristics. Differences between the Buchner funnel test and an operating filter include: the sludge is top fed to the Buchner funnel filtering medium and the test filter medium is much tighter. Therefore, it is not possible to accurately predict the solids concentration of the filtrate, nor cake release characteristics.

The filter leaf test, however, permits an accurate prediction of the operation of a full-scale filter. The filter leaf test employs the use of a test leaf over which is fitted a filtering medium identical to that which will be used on the full-scale filter. The procedure for conducting filter leaf tests described by Eckenfelder and O'Connor [15] is typically to:

1. Condition approximately 2 liters of sludge for filtration. The sludge should be thickened to a minimum concentration of 2 percent or to that anticipated for the full-scale application.

2. Apply desired vacuum to filter leaf and immerse in sample 1½ min (maintain sample mixed). The test leaf normally is inserted upside down in a

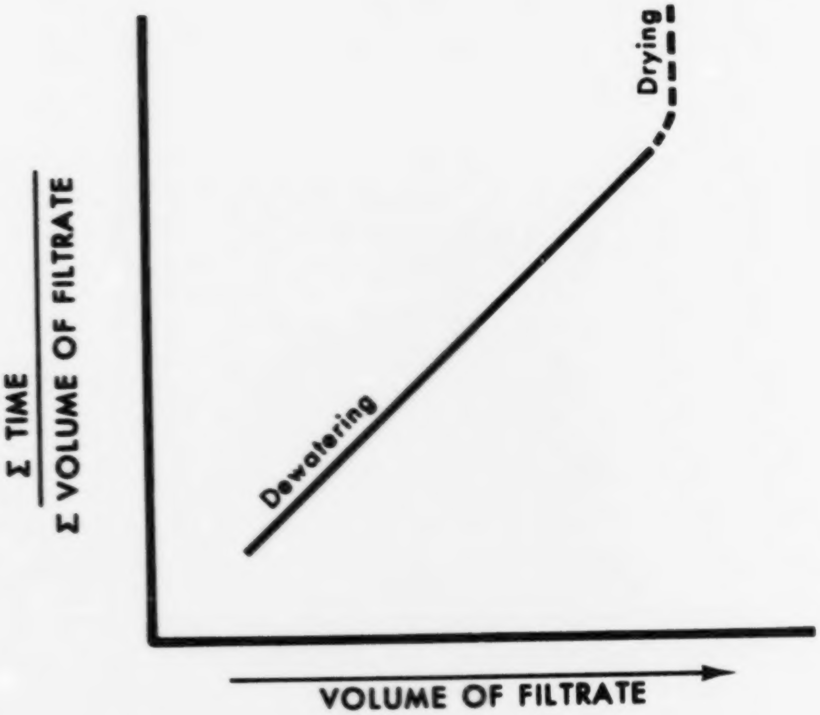

FIGURE 7-5. Typical Buchner funnel test plot.

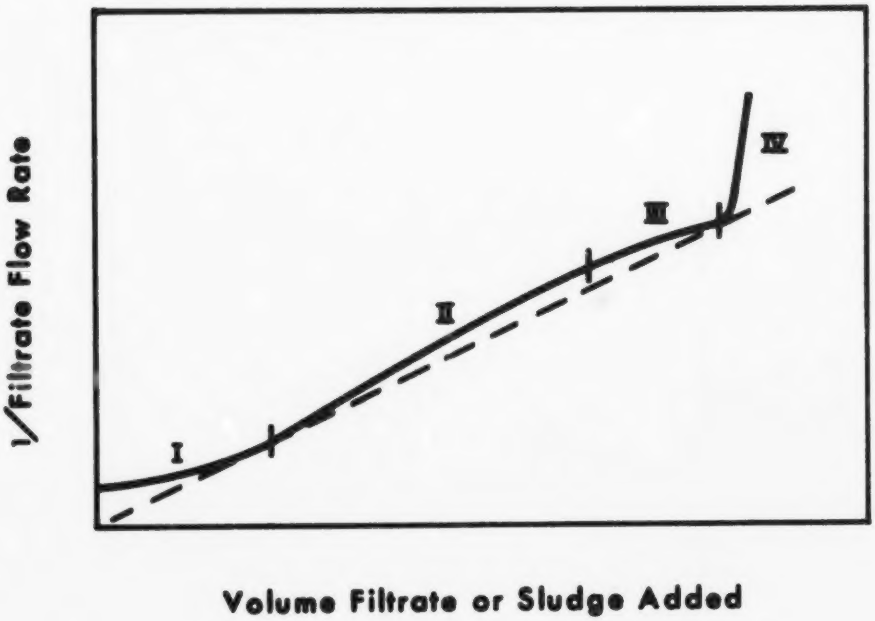

FIGURE 7-6. Instantaneous filtrate flow rate.

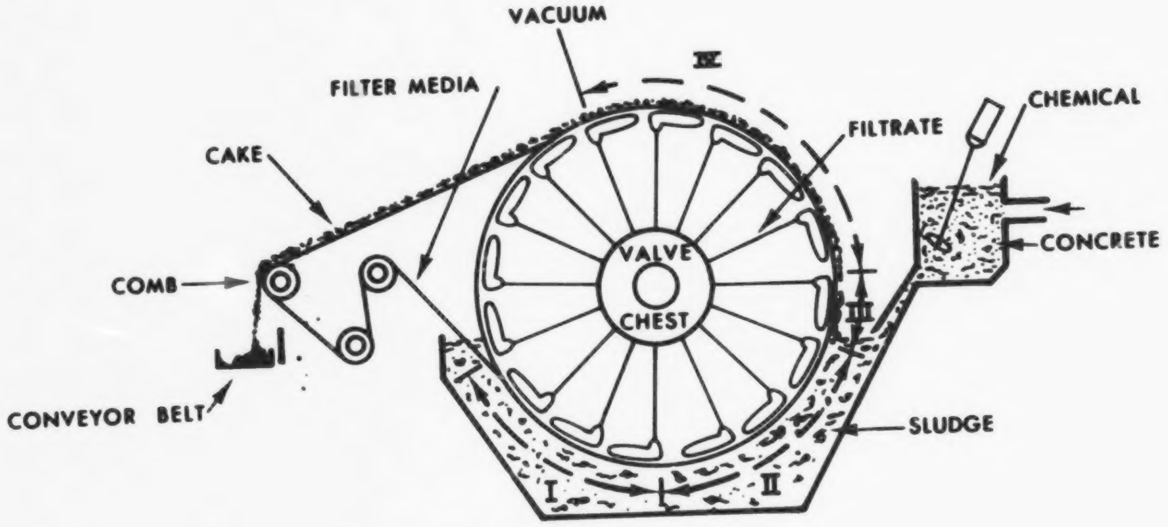
VACUUM
IV
FILTER MEDIA
CHEMICAL
FILTRATE
CAKE
CONCRETE
VALVE
CHEST
COMB
III
CONVEYOR BELT
SLUDGE
I
II

FIGURE 7-7. Cake processing phases rotary vacuum filter.

representative slurry to simulate the cake formation zone of the drum filter. This portion of the cycle is cake formation.

3. Bring leaf to vertical position and dry under vacuum for 3 min (or other predetermined time). This is the cake draining and drying part of the cycle.

4. Blow off cake for 1½ min (this gives a total drum cycle of 6 min). To discharge the cake, the leaf is disconnected and air applied (pressure not exceeding 2 psi).

5. Dry and weigh cake to determine percentage moisture. The filter rate (Y) in $\text{lb/ft}^2/\text{hr}$ is computed:

$$Y = \frac{\text{dry weight sludge, (gm)} \times \text{cycles/hr}}{453.6 \times \text{test leaf area (ft}^2\text{)}}$$

The test can easily be modified for other cycle times and discharge mechanisms. Filter leafs are readily available from filter manufacturers and include instructions. It may be necessary to adjust the above result by a factor to compensate for partial medium blinding over a long period of operation and scale up. Although the filter leaf test is a simple one, there are some precautions which should be observed to insure accurate results:

- Representative sludge samples must be used.
- Several (5 to 10) tests should be run to monitor filter medium blinding.
- The test sample must be agitated to insure that it is homogeneous.
- The test filter vacuum must be regulated so that it does not vary during the test and so that it is the same as proposed for use in full-scale operation.

Normally, the moisture content of the filter leaf test cakes is plotted as a function of a correlating factor on rectangular coordinates. The equation for the correlating factor is:

$$\frac{\text{cfm}}{\text{sq ft}} \times t_d \times \frac{P}{W'}$$

where

cfm/sq ft = Air flow through the cake per unit area of filtering surface

t_d = Dry time, usually expressed in minutes

P = The vacuum differential, psig

W′ = Weight of the dry cake solids in lb per sq ft for a given cake thickness

A decreasing moisture correlation indicates that, as the air rate through the cake per unit of filtering area is increased, or as the vacuum differential or length of the drying time is increased, the moisture content decreases. Conversely, if the cake thickness and subsequently the cake weight (W′) is increased, the moisture content increases. Knowing the percentage of available drying time of the filter cycle and using the design information (the proper cake thickness for a given type of filter, the vacuum level, and air rate through the cake), it is possible to predict for each cycle time the discharged filter cake moisture contents expected from the full-scale filter.

Discussion

Vacuum filtration of wastewater sludge is governed by the media's opening and the size distribution of solid particles in the sludge. Raw primary sludges mainly contain particles smaller than 100 mesh (0.15 mm). Filter cake formation is accomplished first by a blinding of the media with the larger particles and this is followed by a packing of the pores near the filter media with the fine particles [14]. An effect of elutriation is to remove small particles that pack the pore openings. Chemical conditioning changes the size of sludge particles and eliminates the large number of very small particles. As noted in Chapter 6, polyelectrolytes and inorganic chemicals act differently. Both agglomerate the fine particles, reduce the resistance, and clarify the filtrate. However, polyelectrolytes agglomerate the fine particles and attach them to the larger ones. Iron and lime tend to precipitate a coating on the fine particles, making them larger so that they do not pack into the smaller pore openings. Heat treatment solubilizes many of the fine particles and others are removed in the decant step. Two ways to increase water removal rates in vacuum filtration are by using a coarse filter media and allowing some fine solids to pass the media at the beginning of the filter cycle, and chemical conditioning. These techniques are illustrated in Figure 7-8 along with the combination of media size and conditioning considerations.

Where chemical conditioning of sludge is employed, coarse filter media is often used to take advantage of the combined effects. Five to 10 percent solids recycle to the treatment plant from the filtrate is common and this seems to be the most economical method to accomplish vacuum filtration where chemical conditioning is used. When coarse filter media is used, the machine piping maximum discharge rate controls the flow at the beginning of the cycle. The machine variables, such as submergence and drum speed, are not very sensitive for coarse media filters. In general, the cake will form until the complete capture phase occurs and then the buildup will be very slow. Increasing the formation time beyond that point does not appreciably change the cake thickness.

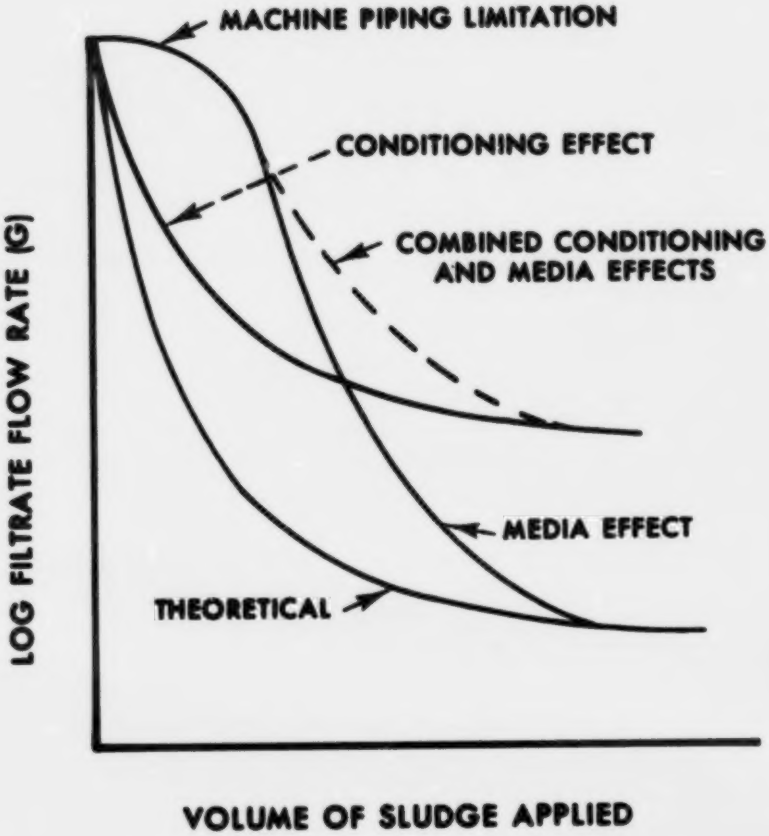

FIGURE 7-8. Media size and conditioning effects on filtrate flow rate.

7.2.2 Process Objectives

As was noted previously, the operator of a vacuum filter strives for maximum solids capture, filter cake yield, and filter cake solids content. He also wants to minimize his costs. The relative importance of these objectives varies with the method of filter cake disposal, and it is not usually possible to accomplish all of them. This necessitates striking a reasonable balance.

Solids Capture

The amount of solids which can be recycled to a wastewater treatment process and not affect its efficiency sometimes needs to be determined experimentally. However, in biological plants, experience indicates that greater than 90 percent total solids capture in the dewatering stage is usually required.

Solids capture is affected by:

- Relative proportion of suspended and dissolved solids in the sludge.
- Sludge characteristics, conditioning, and media.
- Filter drum washing.

The rotary vacuum and horizontal belt filters are capable of producing the highest filtrate quality and hence solids capture of the various continuous dewatering alternatives. Solids capture by vacuum filters may range from 85 to 99.5 percent depending on the type of filter media, chemical conditioning, and solids concentration in the applied sludge. A material balance over the projected dewatering procedure is an essential feature of effective design.

Cake Yield

Units of expression are pounds of dry total sludge solids discharged from the media per hour, per square foot of filter area. It is important to note the inclusion of the term sludge solids in the definition of filter yield. When large percentages of lime, ferric chloride, or ash are used for conditioning and largely end up in the cake, correction of cake yields and solids contents must be made to maintain the validity of these basic terms. Sufficient filter area must be provided so that the sludge solids removal rate necessary to prevent excessive solids accumulation in the plant can be maintained. Since the dewatering step is one of the two directly measurable and legitimate points for removal of solids from the total plant process, maintenance of continuous and adequate solids removal is absolutely essential to efficient system operation. The filter area provided for in design should be for the peak sludge removal rate required plus a 5 to 15 percent area allowance for maintenance downtime. Cake yield is affected by essentially the same parameters of operation as is solids capture. Rotary vacuum filter cake yields may vary from 2 to 15 lb/hr/ft^2, but a yield of less than 3.5 is normally an indication of some problem in sludge process design or operation. Raw primary yields of 7 to 15 lb/hr/ft^2, digested primary yields of 4 to 7 lb/hr/ft^2, and mixed digested yields of 3.5 to 5 lb/hr/ft^2 are typical. The effect of feed solids content on yield is shown in Figure 7-9. The dependence of filter cake yield on feed solids content graphically illustrates the benefit of thickening sludges prior to dewatering by vacuum filtration. A relationship between labor costs and filter yield has been developed by Bennett [2] for an assumed situation and is shown in Figure 7-10. At yields below 4 lb/hr/ft^2, labor costs increase rapidly.

Cake Solids Content

Cake solids content is affected by the sludge type, sludge solids concentration, mode of conditioning, and machine operation. The interdependence of cake solids level, feed solids content, and type of sludge is illustrated in Figure 7-11. As with filter yield, thickening of the feed sludge develops a drier filter cake. The data from McCarty [16] as plotted in Figure 7-11 is a compilation of information from various plants and is an average for the different types of sludge. The other curves are for the four types of sludge processed at the Metro Denver plant. It should be noted that the sludges encountered at Denver are unusually difficult to process. The Denver case history will be discussed later in Chapter 10.

Cake solids concentration is also a very important consideration when incineration or trucking of cake to land disposal are contemplated. As will be noted in Chapter 10, cake solids concentration is only one of the significant parameters in gauging the effectiveness of various preincineration sludge processing systems.

7.2.3 Types of Rotary Vacuum Filters

The three principal types of rotary vacuum filters are shown.

Type	Covering Used	Discharge Mechanism
Drum	Cloth	Blowback section and doctor blade
Coil	Stainless steel spring	Coil layer separation and tines
Belt	Cloth, infrequently metal	Small diameter roll, or flapper, or doctor blade

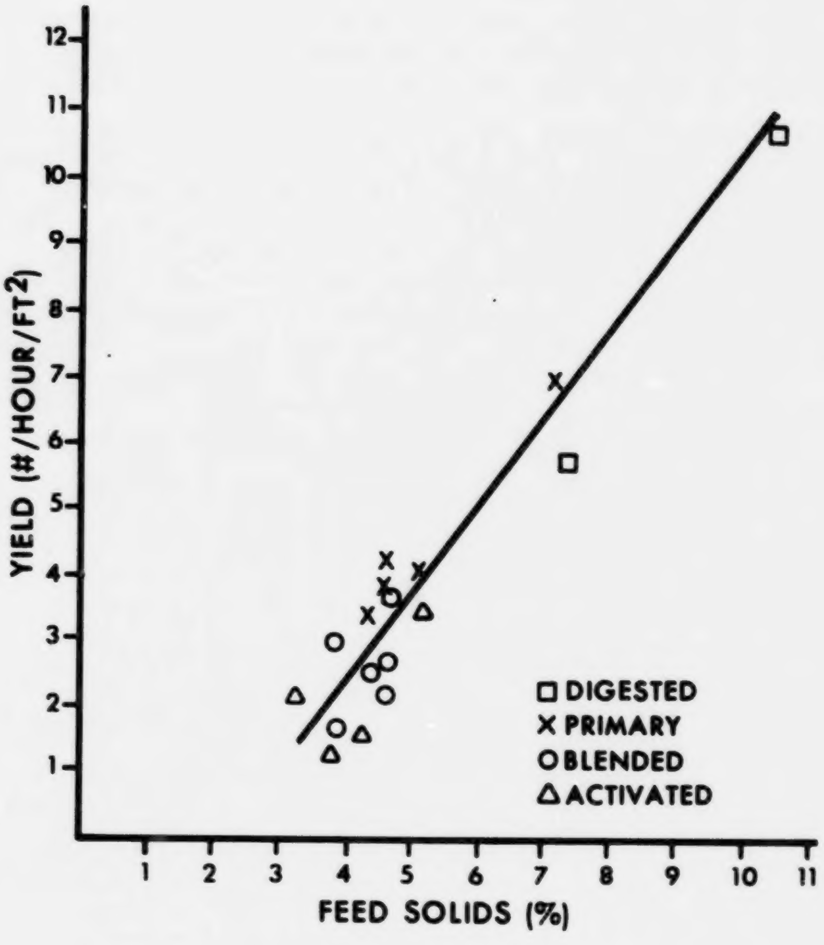

FIGURE 7-9. Yield as a function of feed solids [2].

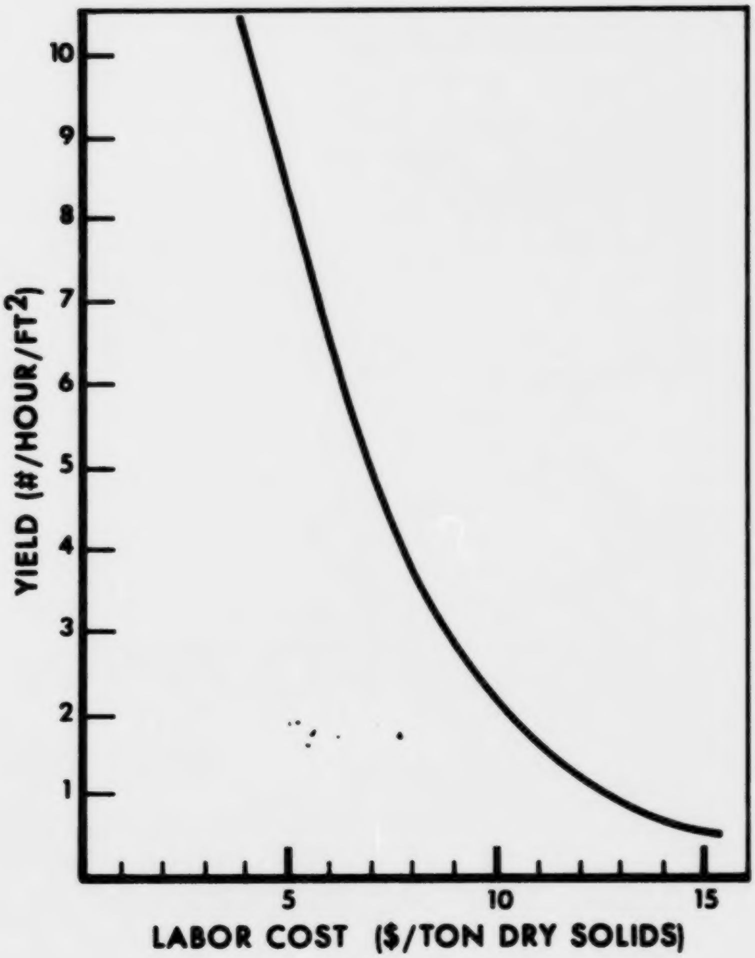

FIGURE 7-10. Vacuum filtration operational labor costs as function of yield [2].

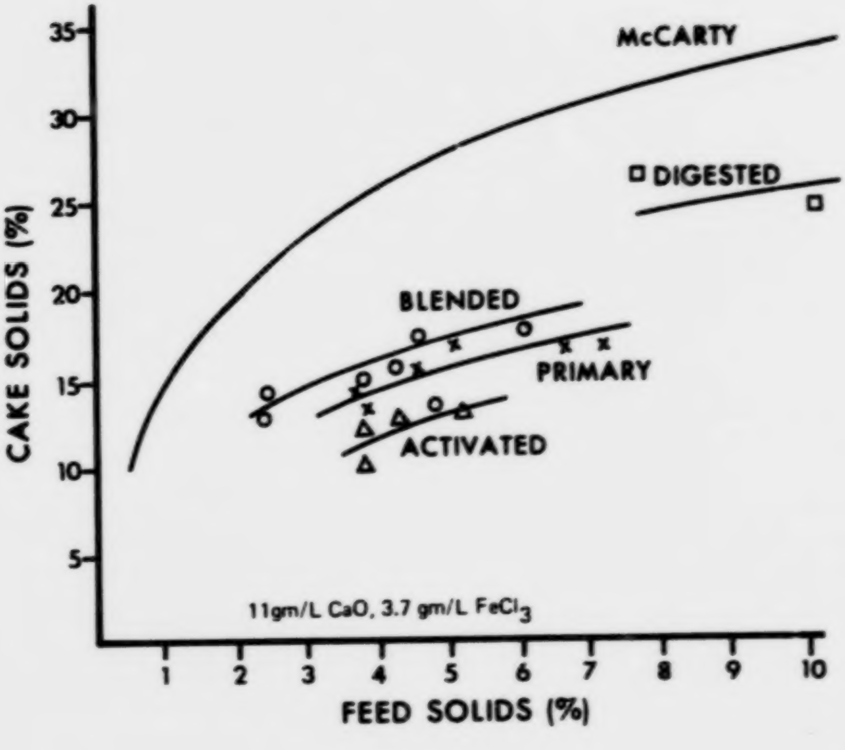

FIGURE 7-11. Cake solids as a function of feed solids for different sludges [2].

The filters differ primarily in the type covering used and the cake discharge mechanism employed. The drum filter (see Figure 7-1) also differs from the other two in that the cloth covering does not leave the drum but is washed in place, when necessary. The design of the drum filter provides considerable latitude in the amount of cycle time devoted to cake formation, washing, and dewatering; while it minimizes inactive time. The drum filter was the original type unit employed in municipal wastewater plants. Problems with frequent washings of the drum cloth when large doses of lime were being used for conditioning have been essentially eliminated with improved filter media and the use of polymers.

A variation of the conventional drum filter is the top feed drum filter. In this case, sludge is fed to the vacuum filter through a hopper located above the filter. The city of Milwaukee, Wisconsin, is currently planning plant scale evaluation of this type of vacuum filter [17]. The potential advantages are that gravity aids in cake formation; capital costs may be lower since the feed hopper is smaller and no sludge agitator and related drive equipment are required; and blinding may be reduced because the gravity formation of the cake allows larger particles to collect first on the cloth providing a straining layer to capture the smaller particles.

The coil type vacuum filter is shown in Figure 7-12 and uses two layers of stainless steel coils arranged in corduroy fashion around the drum. After a dewatering cycle, the two layers of springs leave the drum and are separated from each other so that the cake is lifted off the lower layer of springs and discharged from the upper layer. Cake release is essentially never a problem. The coils are then washed and reapplied to the drum. Figure 7-13 shows a cutaway view of the coil springs. The coil filter has been and is widely used for all types of sludge. However, sludge with particles that are both extremely fine and resistant to flocculation dewater poorly on coil filters.

The belt type filter is shown in Figure 7-14. Media on the belt filter leaves the drum surface at the end of the drying zone and passes over a small diameter discharge roll to facilitate cake discharge. Washing of the media next occurs before it returns to the drum and to the vat for another cycle. This type filter normally has a small diameter curved bar between the point where the belt leaves the drum and the discharge roll. This bar primarily aids in maintaining belt dimensional stability. In practice it is frequently used to insure adequate cake discharge. Figure 7-15 shows the demooning bar set at a maximum angle of projection to help break the cake free from the belt. Remedial measures are frequently required to obtain operable cake releases from belt filters. This is particularly true at plants which have greasy or sticky sludges due to high activated sludge content. A summation of locations and type of remedial action required to obtain cake release from belt type filters appears in Table 7-2.

Rotary vacuum filters are normally supplied with essential auxiliary equipment. A complete system is shown in Figure 7-16. Principal auxiliary equipment includes a

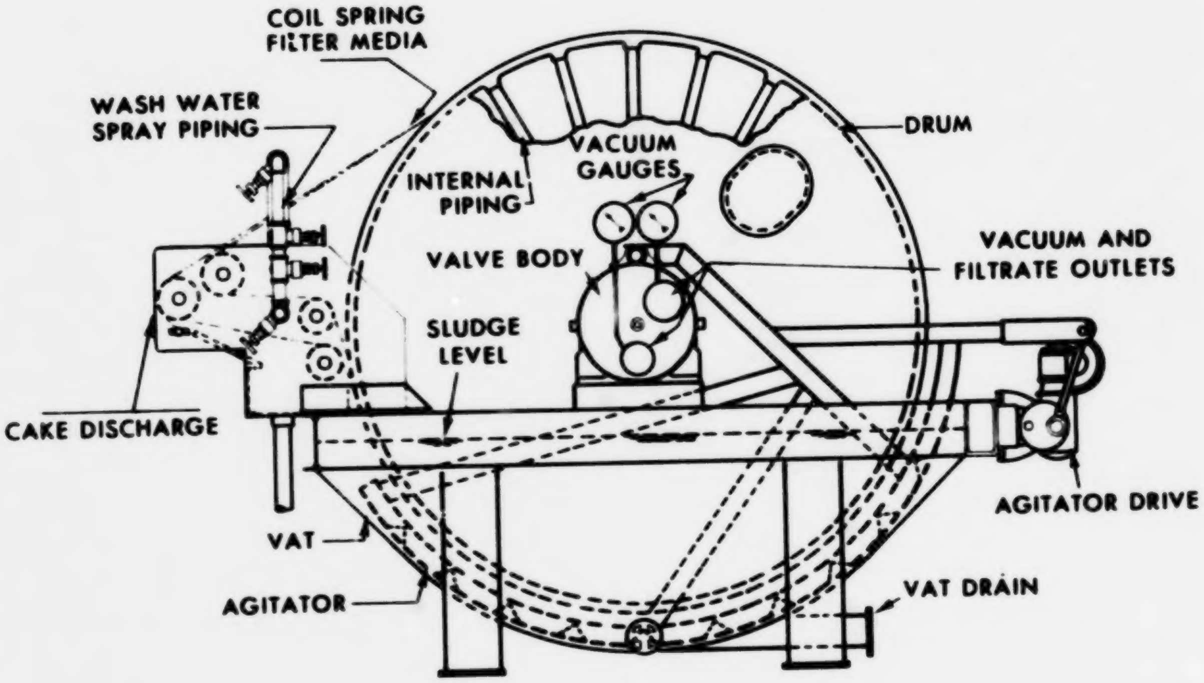
COIL SPRING FILTER MEDIA
WASH WATER SPRAY PIPING
DRUM
VACUUM GAUGES
INTERNAL PIPING
VALVE BODY
VACUUM AND FILTRATE OUTLETS
SLUDGE LEVEL
CAKE DISCHARGE
AGITATOR DRIVE
VAT
AGITATOR
VAT DRAIN

FIGURE 7-12. Cross section of a coil filter.

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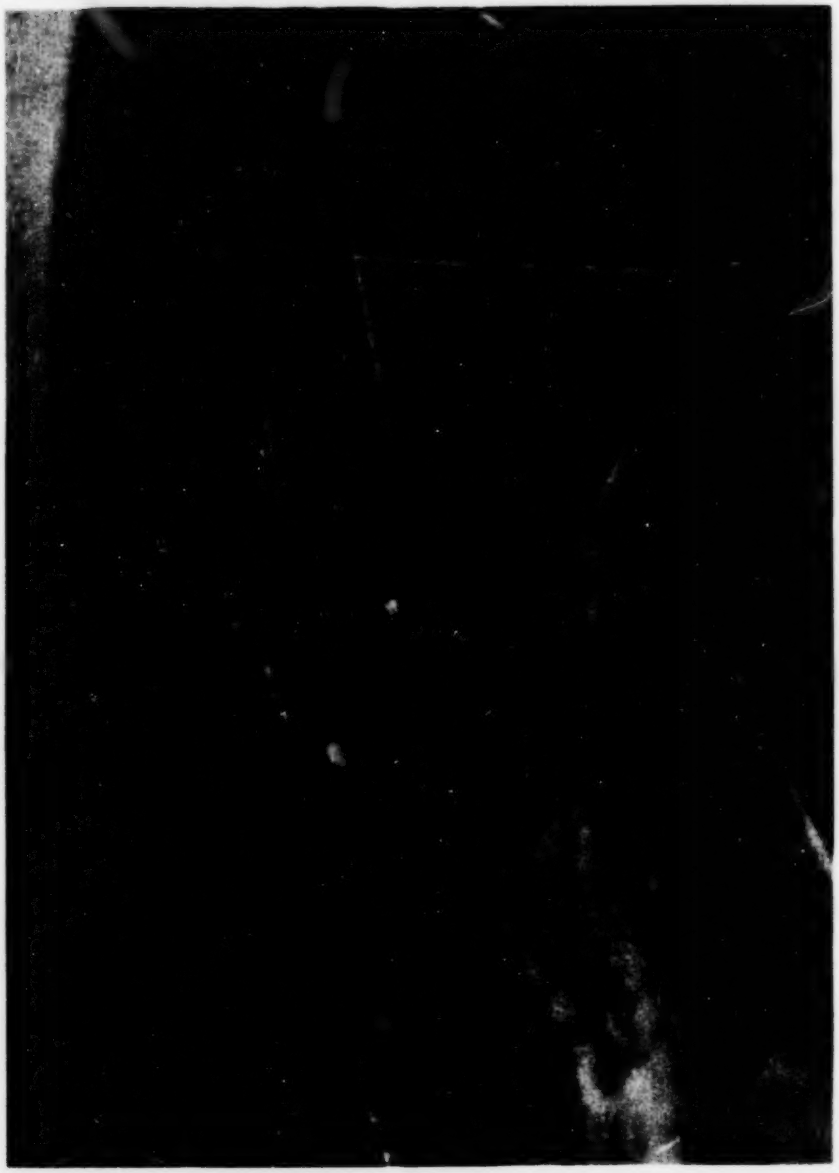


FIGURE 7-13. Cutaway view of coil springs.

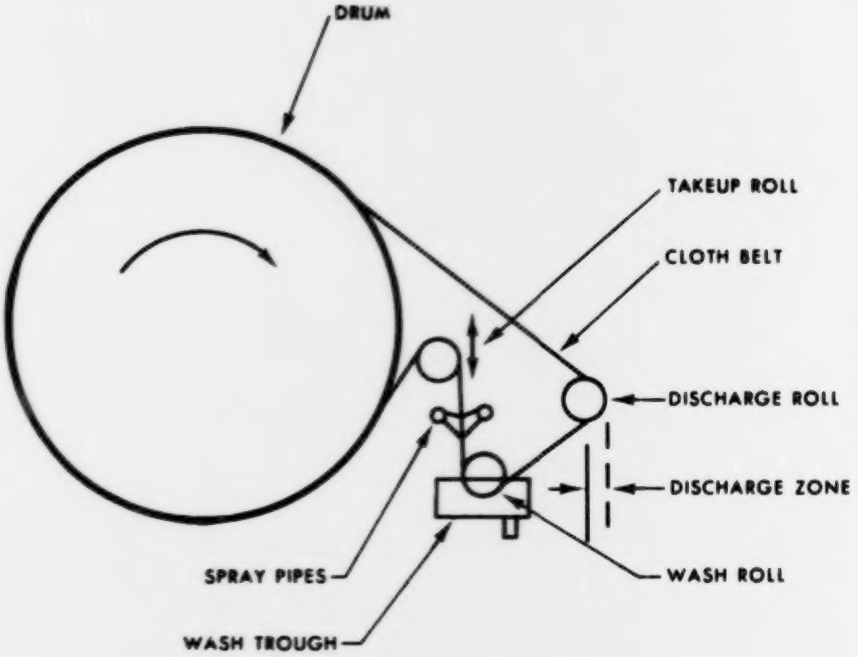

DRUM
TAKEUP ROLL
CLOTH BELT
DISCHARGE ROLL
DISCHARGE ZONE
WASH ROLL
SPRAY PIPES
WASH TROUGH

FIGURE 7-14. Cross section of a belt filter.

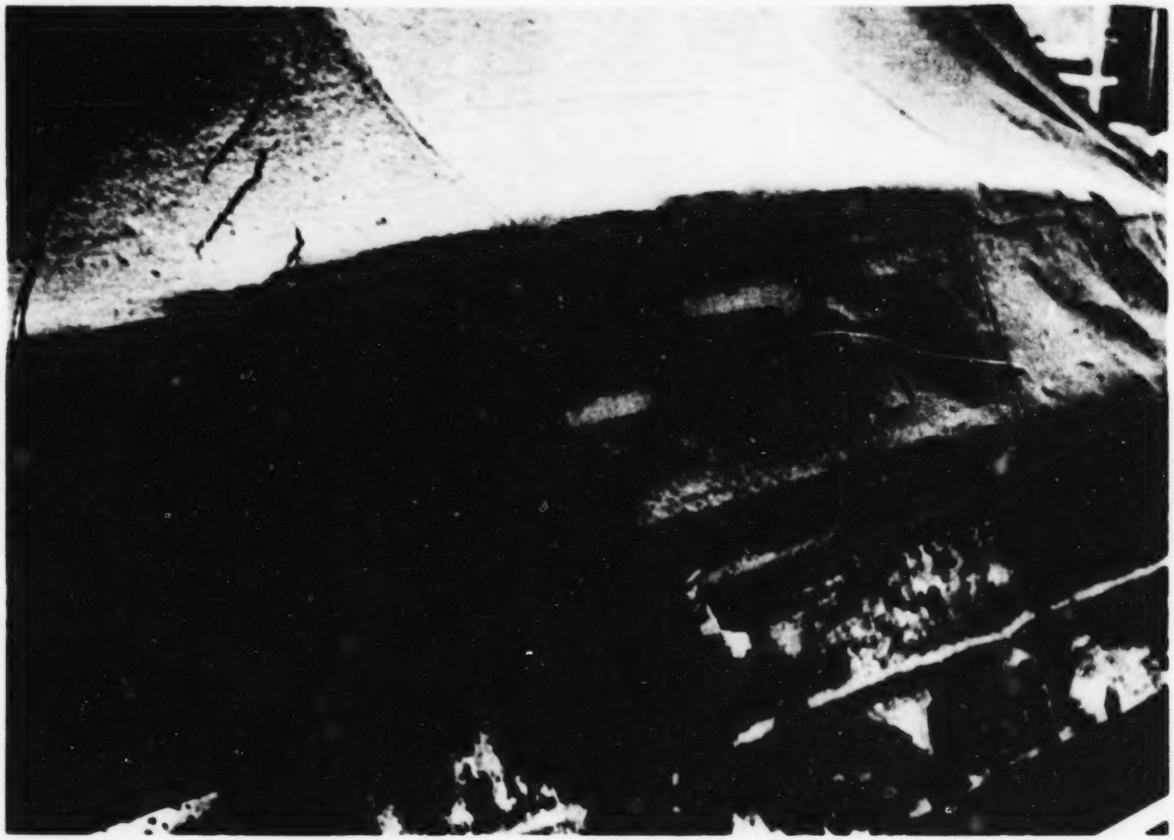


FIGURE 7-15. Cake release of a belt filter.

TABLE 7-2

CAKE RELEASE MEASURES USED ON BELT TYPE FILTERS AT VARIOUS PLANT LOCATIONS

Location	Conditioning System	Remedial Steps Required to Obtain Discharge
Washington, D.C.	$FeCl_3$ and Polymer	Use of excess amount of ferric chloride 3 lb/ton
Baltimore, Md.	Polymer	Plastic doctor blade added
Indianapolis, In.	Ash and Polymer	Inclusion of incinerator ash with sludge
Toronto, Ontario	Polymer or $FeCl_3$/Lime	None successful; belt filters idle
Richmond, Ca.	$FeCl_3$ and Lime	Chemical dose increased
Colorado Spgs., Co.	Heat Treatment	Occasional operator assistance

vacuum pump, filtrate receiver and pump, sludge conditioning apparatus, and a sludge pumping system. Usually one vacuum pump is provided for each vacuum filter although some larger plants use fewer pumps connected to a common header. In early days, reciprocating type dry vacuum pumps were generally specified but wet type vacuum pumps are now almost universally used. The wet type pumps are more easily maintained and provide sufficient vacuum. Wet type pumps utilize seal water and it is essential that a satisfactory water be used. If the water is hard and unstable, it may be necessary to prevent carbonate buildup on the seals through the use of a sequestering agent. Each vacuum filter must be supplied with a vacuum receiver interposed between the filter valve and the vacuum pump. The receiver is usually designed to give a maximum air velocity of 2.5 to 5 ft/min and a minimum air detention time of 2 to 3 minutes [6]. The principal purpose of the receiver is to separate the air from the liquid. Each receiver can be equipped with a vacuum limiting device to admit air flow if the design vacuum is exceeded, a condition which would cause the pump to overload. The receiver also acts as a reservoir for the filtrate pump suction. Filtrate pumps must be sized to carry away the water separated in the vacuum receiver. These are specially designed pumps to operate at very low net positive suction heads and are designed for at least 20- to 22-inch Hg vacuum at the inlet. The discharge head depends on the local conditions.

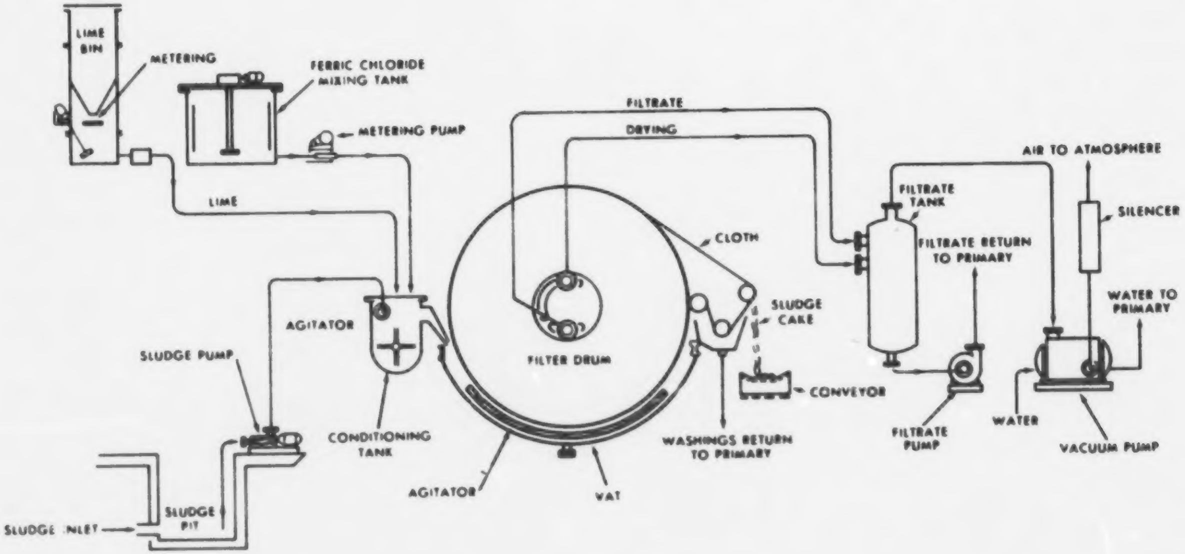

LIME BIN
METERING
FERRIC CHLORIDE MIXING TANK
METERING PUMP
LIME
FILTRATE
DRYING
AIR TO ATMOSPHERE
FILTRATE TANK
SILENCER
CLOTH
FILTRATE RETURN TO PRIMARY
WATER TO PRIMARY
SLUDGE CAKE
AGITATOR
SLUDGE PUMP
FILTER DRUM
CONVEYOR
WATER
CONDITIONING TANK
WASHINGS RETURN TO PRIMARY
FILTRATE PUMP
VACUUM PUMP
AGITATOR
VAT
SLUDGE PIT
SLUDGE INLET

FIGURE 7-16. Rotary vacuum filter system.

Centrifugal style filtrate pumps are common but can become air bound unless they have a balance or equalizing line connecting from a high point of the receiver to the eye of the pump. Nonclogging centrifugal style pumps are used with coil filters or with coarse metal filter media. They permit a somewhat higher solids concentration in the filtrate. Self-priming centrifugal pumps are used most frequently, since they are relatively maintenance free. Self-priming, nonclogging centrifugal pumps are also available. Check valves on the discharge side of the pumps are usually provided to minimize air leakage through the filtrate pump and receiver back to the vacuum pump. Filtrate pumps should be sized to accommodate the entire range of filtrate flow rates likely to be encountered. The fact that the rate of filtrate flow is a function of the mode of conditioning must be recognized in filtrate pump sizing. Polyelectrolytes allow the sludge to drain much more rapidly than do inorganic conditioners. If the filtrate pumps are not sized accordingly, full advantage of the more rapid drainage cannot be realized.

7.2.4 Machine Variables

The operation of a rotary vacuum filter is sensitive to the type of sludge and conditioning procedures. Plant experimentation is the best way to test a potentially improved mode of operation. One major machine variable is the media used and a great many types of filter media are available for the belt and drum filters. Blinding characteristics and chemical conditioning play an important role in media selection. Filter leaf tests should be conducted with the various media as an aid in selecting the optimum one for a specific sludge. The ideal media has the following characteristics [18]:

- It is able to perform the desired liquid/solid separation and give a filtrate of acceptable clarity.
- The filter cake discharges readily from it.
- It is strong enough mechanically to give a long life.
- It is chemically resistant to the materials being handled.
- Its resistance to flow is not too great.
- It does not rapidly blind.

Obviously, some reasonable compromise must be reached between these objectives since all of them cannot be optimized simultaneously. The Chicago Sanitary District has the largest vacuum filter installation in the U.S. and has evaluated many different types of filter media, reported by Shedden [19]. Table 7-3 summarizes their conclusions and

TABLE 7-3

EVALUATION OF ALTERNATE FILTER MEDIA AT THE CHICAGO SANITARY DISTRICT [19]

	Material												
Evaluation Criteria	Wool 12 oz	Wool 13 oz	Wool 14-15 oz	Wool 16 oz	Wool Treated 14-15 oz	Vinyon	Nylon 100%	Nylon 90% Saran 10%	Nylon 25% Wool 75%	Dynel 50% Wool 50%	Orlon 100%	Dynel 100%	Dacron 100%
Installation	Difficult	Difficult	Easy	Difficult	Easy	Very Difficult	Very Difficult	Very Difficult	Easy	Easy	Difficult	Difficult	Easy
Initial Cake Pickup	Poor	Fair	Very Good	Fair	Very Good	Fair	Fair	Fair	Good	Good	Fair	Poor	Very Good
Cake Production	Fair	Fair	Very Good	Fair	Very Good	Fair	Fair	Fair	Fair	Good	Good	Good	Very Good
Elongation Tendency	Excessive	Excessive	High	High	Some Varied	Excessive	High	High	Some	Some	Some	Some	Very Little
Response to Water Wash	Good	Good	Good	Good	Very Good	Fair	Fair	Fair	Good	Good	Good	Good	Very Good
Response to Detergent Wash	Good	Good	Good	Good	Very Good	Poor	Fair	Fair	Good	Good	Fair	Good	Very Good
Response to Acid Wash	Good	Good	Good	Good	Very Good	Poor	Very Poor	Poor	Poor	Good	Poor	Good	Very Good
Resistance to Scraper Abrasion	Poor	Fair	Good	Good	Good	Poor	Good	Good	Good	Good	Fair	Fair	Very Good
Useful Life (approx. hrs)	1900	2400	2700	1600	3300	1000	800 to 7600	3000	3300	3700	2000	Over 4000	15,000
Economic Rating	-	-	4	-	3	-	-	-	-	-	-	2	1

indicates that Dacron, a polyester, was the most suitable to their use. Other treatment plants have found polypropylenes to be satisfactory. Polyethylenes tend to stretch when wet and require constant operator vigilance of belt tension. Minneapolis-St. Paul has reported a life of 12,400 hours for a Saran medium, according to Simpson and Sutton [20]. Monofilament fabrics are the most resistant to blinding and have been used almost exclusively in recent installations of drum or belt filters.

Up to a point, filter yield increases as the vacuum is increased. Because of the compressible nature of wastewater sludges, there is some question whether operating vacuums greater than 15 inches of mercury are justifiable. Only slight increases in yield are normally experienced beyond this level [21]. The cost of a greater filter area must be balanced against the higher power costs for higher vacuums. An increase from 15 to 20 inches of vacuum is reported by Schepman and Cornell [22] to have provided about 10 percent greater yield in three full-scale installations.

Increasing the drum submergence rate increases the form cycle time and usually results in an increased yield, and thicker but wetter cake. In general, the maximum submergence used on a sludge filter is 25 percent, although higher submergences are possible. The submergence is usually kept between 15 and 25 percent which gives a long drying time and keeps the cake moisture content at a minimum.

Slowing the drum speed increases the filter cycle time and produces a drier cake, but the filter's productivity is decreased.

Proper agitation of the sludge during and after chemical conditioning is important. The evaluation of this parameter requires variable speed mixing equipment for both the chemical conditioning tanks and the vacuum filter pan. After chemical conditioning, the sludge must be handled as gently as is practical. Only enough agitation should be applied in the filter pan to prevent solids classification and keep the solids in suspension. Because sludge viscosities vary, optimum control requires variable speed pan agitation equipment.

7.2.5 Rotary Vacuum Filter Costs

Capital costs may range from $100 to $300 per square foot depending on unit size, type media, and auxiliary equipment [23]. Operation and maintenance costs vary widely according to plant size, pretreatment procedures, and product quality requirements, but $5 to $20 per ton is representative. An approximate breakdown in elements of vacuum filter O/M costs is provided by Simpson and Sutton [24].

	Percent
Labor and direct supervision	39
Chemicals and supplies	37

	Percent
Electric power	8
Maintenance	16

7.2.6 Typical Rotary Vacuum Filter Results

Table 7-4 presents representative data for sludges conditioned with ferric chloride and lime.

TABLE 7-4

TYPICAL ROTARY VACUUM FILTER RESULTS FOR SLUDGE CONDITIONED WITH INORGANIC CHEMICALS

	Chemical Dose (lb/ton)		Yield	Cake
Type Sludge	Ferric Chloride	Lime	(lb/hr/ft^2)	Solids (%)
Raw Primary	1-2	6-8	6-8	25-38
Anaerobically Digested Primary	1-3	6-10	5-8	25-32
Primary + Humus	1-2	6-8	4-6	20-30
Primary + Air Activated	2-4	7-10	4-5	16-25
Primary + Oxygen Activated	2-3	6-8	5-6	20-28
Digested Primary and Air Activated	4-6	6-19	4-5	14-22

The data in this table are never a substitute for actual lab or pilot tests' results for a particular sludge. Typical data for sludges conditioned with polyelectrolytes are shown in Table 7-5.

The price of polymer per pound can vary considerably. Therefore, these data are presented only to illustrate rough ranges, and a determination of accurate and meaningful unit process costs must be considered as an integral part of a particular system.

TABLE 7-5

TYPICAL ROTARY VACUUM FILTER RESULTS FOR POLYELECTROLYTE CONDITIONED SLUDGES

Type Sludge	Chemical Cost ($/ton)	Yield (lb/hr/ft^2)	Cake Solids (%)
Raw Primary	1-2	8-10	25-38
Anaerobically Digested Primary	2-5	7-8	25-32
Primary + Humus	3-6	4-6	20-30
Primary + Air Activated	5-12	4-5	16-25
Primary + Oxygen Activated	5-10	4-6	20-28
Anaerobically Digested Primary and Air Activated	6-15	3.5-6	14-22

7.2.7 Summation

Rotary vacuum filtration can be and is, in most cases, an effective and efficient dewatering method. It has been misapplied in some cases in the past. Improper selection of media, failure to thicken the feed sludge, the cake release problem on belt filters, and lack of proper sludge conditioning have generally been the causes of failures. In some cases, as will be discussed in Chapter 10, plant systems have not been designed with the proper sequence of unit processes. This makes efficient dewatering very difficult.

7.3 Centrifugal Dewatering

Centrifuges of various types have been employed for solid-liquid separation processes in agriculture and industry for at least 50 years. For almost 25 years the continuous solid bowl conveyor type centrifuge has been used for dewatering municipal sludges. Objectives of centrifugal sludge dewatering are the same as for rotary vacuum filtration.

7.3.1 Theory of Centrifugal Dewatering

Solids Removal and Conveyance

The centrifuge uses centrifugal force to speed up the sedimentation rate of sludge solid particles. Figure 7-17 shows a continuous solid bowl dewatering centrifuge. The two principal elements of this centrifuge are the rotating bowl which is the settling vessel and the conveyor discharge of settled solids. The bowl has adjustable overflow weirs at its larger end for discharge of clarified effluent (centrate) and solids discharge ports on the opposite end for discharging dewatered sludge cakes. As the bowl rotates, centrifugal force causes the slurry to form an annular pool, the depth of which is determined by the adjustment of the effluent weirs. A portion of the bowl is of reduced diameter so that it is not submerged in the pool and thus forms a drainage deck for dewatering the solids as they are conveyed across it. Feed enters through a stationary supply pipe and passes through the conveyor hub into the bowl itself. As the solids settle out in the bowl, they are picked up by the conveyor scroll and continuously carried along to the solids outlets. Clear effluent at the same time continuously overflows the effluent weirs. Flocculants are normally injected into the pool.

It is extremely important to note that there are two operating zones in the horizontal bowl conveyor centrifuge; the submerged pool and the drainage deck. Early theoretical consideration of centrifugal dewatering mechanisms focused primarily on the relationship between the centrifuge and a hypothetical sedimentation basin as affected by the employment of very high "G" forces. The Sigma formula from Perry's *Chemical Engineers Handbook* [25] is normally employed to describe the operation of a continuous, horizontal, helix-type centrifuge and is shown here. This formula shows that the rate of liquid clarification varies with the surface area of the liquid and the level of centrifugal force.

$$\Sigma = \frac{\pi b w^2}{2g} \left(3r_2^2 + r_1^2\right)$$

where

Σ = Sigma centrifuge capacity factor, ft^2 (Theoretical area of gravity settling tank of equivalent sedimentation characteristics to centrifuges)

b = Length of cylindrical bowl, ft

w = Rate of rotation, radians/second

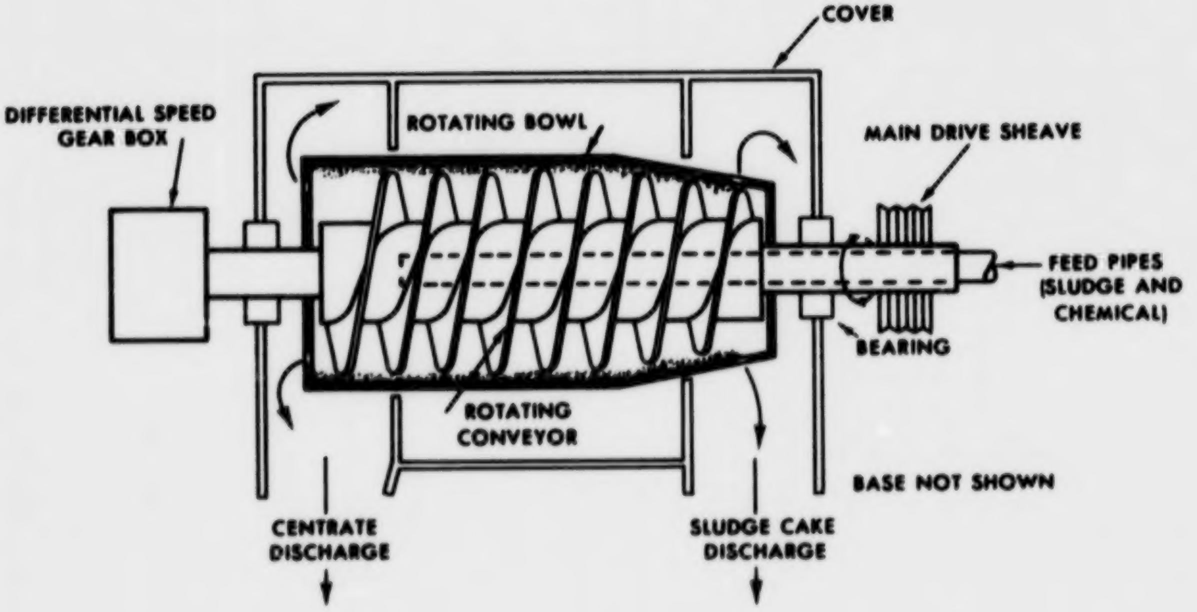

COVER
DIFFERENTIAL SPEED GEAR BOX
ROTATING BOWL
MAIN DRIVE SHEAVE
FEED PIPES (SLUDGE AND CHEMICAL)
BEARING
ROTATING CONVEYOR
BASE NOT SHOWN
CENTRATE DISCHARGE
SLUDGE CAKE DISCHARGE

FIGURE 7-17. Continuous countercurrent solid bowl conveyor discharge centrifuge.

r_2 = Radius of inner bowl wall, ft

r_1 = Radius of retained liquid surface, ft

g = Gravitational constant, ft/sec^2

Sigma and other theoretical relationships based on easily measured machine dimensions are useful tools when employed by the centrifuge designer for estimating scale-up relationships in geometrically similar machines. Note that a factor of two, missing from the formula in the referenced handbook, has been included here. Unfortunately, the widespread use of the Sigma formula in the literature and its recent publication in Perry's *Chemical Engineers Handbook* have lead to some centrifuge specifications based only on square feet of Sigma. White pointed out that this can be a serious error, because it theoretically suggests that ample clarifying ability is the only requirement for scale-up and desired performance. No consideration has been given of the solids conveying aspects [26]. The centrifugal force can adversely affect what theory would indicate. This is primarily because a solid bowl conveyor centrifuge not only has to clarify a slurry and settle particles, but it must also accomplish the secondary function of conveying the solids. Thus, while increasing the centrifugal force and lowering the depth of the liquid in the bowl will theoretically increase its clarification ability, it may actually harm clarification in the centrifuge as discussed by White [27]. The design engineer must fully take into account the existence and impact of the drainage deck or beach zone since significant drainage or dewatering of sludge solids occurs here.

Factors Affecting Centrifugal Dewatering

Sludge characteristics which affect centrifuge performance are essentially the same as those listed for rotary vacuum filters. In general, those sludges which separate most readily and concentrate to greatest thickness by plain sedimentation are those which dewater most efficiently in centrifuges.

Machine variables of importance are:

- Bowl design

    Length/Diameter Ratio

    Bowl Angle

    Flow Pattern

- Bowl Speed

- Pool volume
- Conveyor design
- Relative conveyor speed
- Sludge feed rate

The settling time and surface area can be increased for a given diameter bowl by increasing the length/diameter ratio. Although the detention time is increased by an increase in bowl diameter, lower centrifugal forces result because of mechanical limitations. Length/diameter ratios of 2.5 to 3.5 are customarily employed.

The designer can increase the length of the clarifying zone of the bowl by making the discharge angle of the screw conveyor steeper. Centrifugal forces can also be increased. The effect of these two variables on the settled sludge on the conveyor's incline section is illustrated in Figure 7-18.

The slippage force (g) = $\sin\alpha$ while the centrifugal force (G) = $1.42 \times 10^{-5}\ (\text{rpm})^2$. Albertson and Guidi noted that as either the angle of the beach increases or the centrifugal force level increases, the forces driving the settled sludge back down into the pool are increased proportionately [28]. Thus, although the settling rate increases in proportion to the centrifugal force level, the forces rejecting the material into the bowl of the centrifuge are also proportionately increasing and preventing its discharge. Eventual overflow of these solids with the effluent could result.

The flow pattern in the machine may be based on a countercurrent flow of liquids and solids (as shown in Figure 7-17) or a concurrent flow as is discussed later in this chapter. The primary operating variables are bowl speed and pool volume. While increasing the bowl speed increases the centrifugal forces and favors increased clarification, the settled solids become more difficult to discharge. Excessive bowl speed tends to lock the bowl and conveyor together and increases abrasion.

Pool depth affects both clarification and cake dryness. Lowering the pool exposes more drainage deck area, increases the dewatering time, and produces a drier cake. Within limits, increasing pool depth increases clarification by increasing detention time. However, just as in plain sedimentation, too great a depth prevents a particle from reaching the sediment zone prior to being discharged in the effluent. At too shallow a depth, the moving conveyor tends to redisperse settled solids.

Conveyor speeds normally are designed or adjusted to a minimum turbulence inside the pool while still providing sufficient conveying capacity. Low speeds also reduce the rate of wear on the conveyor blades when poorly degritted sludges are handled. Increasing the conveyor speed sometimes produces drier solids because the fines are washed from the cake.

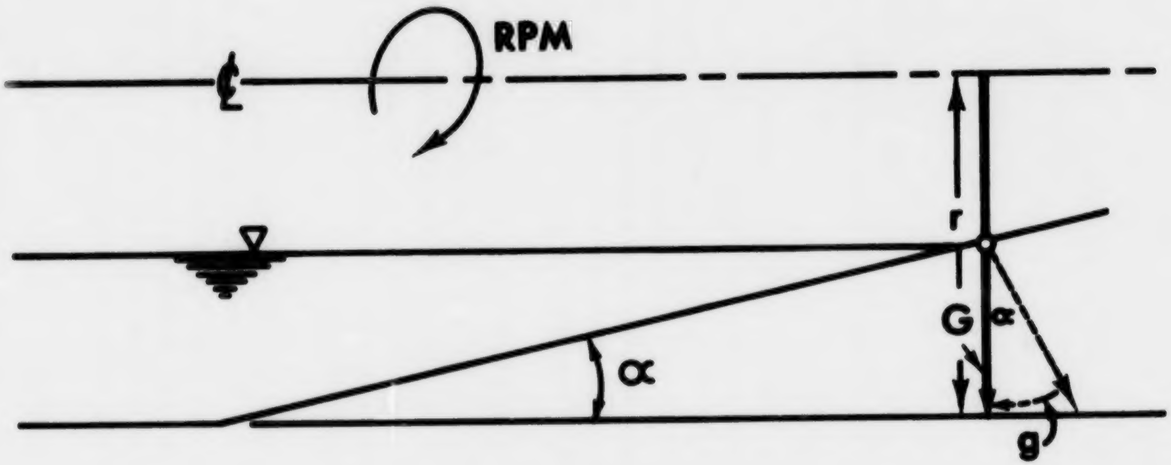


FIGURE 7-18. Effect of bowl angle and centrifugal force on sludge solids in drainage zone.

The sludge feed rate is clearly one of the more important variables. It affects both clarity and sludge cake dryness. The handling of a larger volume of sludge per unit of time in a given bowl means less retention time and a decrease in solids recovery. It also usually results in drier solids in the cake because of the higher loss of fines to the centrate.

Test Procedures

Successful application of continuous, solid bowl, conveyor centrifuges requires a consideration of numerous factors. Proper scale-up is the major factor, and to obtain predictable results, values must be available for the following variables [27]:

- Wet cake discharge rate.
- Solids dewatering time under centrifugal force.
- Conveying torque for cake solids.
- Liquid clarifying ability.
- Resistance to abrasion from slurry solids.
- Stability of centrifuge feed.
- Physical nature of solids being handled.
- Permissible chemical flocculant dosages.

Tests for scale-up can be accomplished with either laboratory or pilot tests. Pilot tests should be conducted with a continuous, small pilot centrifuge geometrically similar to that proposed for full-scale use. Although laboratory tests may be more convenient, they provide little meaningful scale-up data. An evaluation of the relative effects of various flocculants and the potential tradeoffs between high centrifugal force and little flocculant addition and lower force with more flocculants is possible in the laboratory. The only satisfactory method of accurately predicting the performance of a full-scale unit is the operation of a pilot unit on the sludge involved. These pilot units are readily available from manufacturers, who have developed scale-up factors for their pilot equipment. These factors have proven to provide accurate predictions of full-scale performance. These scale-up procedures are considered proprietary and are not generally available. Vesilind [29] presents one approach to scale-up procedures. Since either the hydraulic or the solids capacity may be limiting, scale-up is made for both and the full-scale unit selection is then based on whichever is the governing factor. The scale-up factors are developed in the following equations, where machine 2 is the full-scale unit and machine 1 is the pilot unit. The hydraulic capacity can be estimated as:

$$(Q_L)_2 = \left(\frac{Q_L}{\Sigma}\right)_1 (\Sigma)_2$$

in which

$$\Sigma = \frac{V\omega^2}{g \ln \left(\frac{r_2}{r_1}\right)}$$

The solids handling capacity can be estimated as:

$$(Q_S)_2 = \left(\frac{Q_S}{\beta}\right)_1 (\beta)_2$$

in which

$$\beta = \pi \Delta \omega SDNP$$

The nomenclature used in the preceding equations is:

Q_L = Liquid flow rate

V = Volume in centrifuge occupied by slurry

g = Gravitational constant

r_2 = Radius of bowl wall from center line

r_1 = Radius of slurry surface from center line

Q_S = Solids throughout

β = Differential speed between bowl and conveyor

S = Scroll blades separation

D = Cylinder diameter

N = Number of leads on scroll

P = Pool depth

7.3.2 Types of Centrifuges

Countercurrent

The solid bowl countercurrent centrifuge as discussed earlier is the most widely used type for dewatering of wastewater sludge in the United States. The centrifuge assembly, as has been shown in Figure 7-17, consists of a rotating unit comprising a bowl and conveyor joined through a planetary gear system, designed to rotate the bowl and the conveyor at slightly different speeds. The solid cylindrical-conical bowl, or shell, is supported between two sets of bearings and includes a conical section at one end. This section forms the dewatering beach over which the helical conveyor screw pushes the sludge solids to outlet ports and then to a sludge cake discharge hopper. The opposite end of the bowl is fitted with an adjustable outlet weir plate to regulate the level of the sludge pool in the bowl. This plate also discharges the centrate through outlet ports either by gravity or by a centrate pump attached to the shaft at one end of the bowl. Sludge slurry enters the rotating bowl through a stationary feed pipe extending into the hollow shaft of the rotating bowl. The sludge feed enters a baffled, abrasion-protected chamber for acceleration before discharge through the feed ports of the rotating conveyor hub into the sludge pool in the rotating bowl. The sludge pool takes the form of a concentric annular ring of liquid sludge on the inner wall of the bowl. Separate motor sheaves or a variable speed drive can be used for adjusting the bowl speed for optimum performance.

Bowls and conveyors can be constructed from a large variety of metals and alloys to suit special applications. For dewatering of wastewater sludges, mild steel or stainless steel normally has been used. Because of the abrasive nature of many sludges, hardfacing materials are applied to the leading edges and tips of the conveyor blades, the discharge ports, and other wearing surfaces. Such wearing surfaces may be replaced by welding when required.

Continuous Concurrent Flow Solid Bowl Conveyor Type Centrifuge

Figure 7-19 shows a cross section of a continuous concurrent flow solid bowl conveyor type centrifuge. Incoming sludge is carried by the feed pipe to the end of the bowl opposite the discharge. As a result, settled solids are not disturbed by incoming feed. Solids and liquids pass through the bowl in a smooth parallel flow pattern. Turbulence is substantially reduced. Solids are conveyed over the entire length of the bowl before discharge to provide better compaction, a drier cake, and reduce flocculant demands.

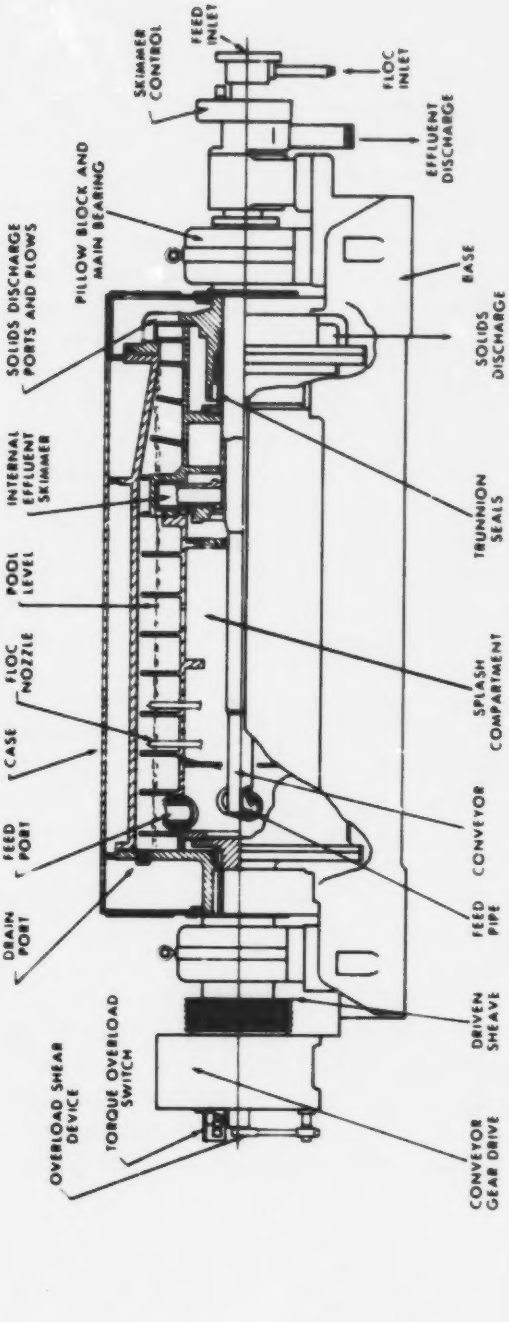

CONVEYOR GEAR DRIVE
OVERLOAD SHEAR DEVICE
TORQUE OVERLOAD SWITCH
DRIVEN SHEAVE
DRAIN PORT
FEED PORT
FEED PIPE
CONVEYOR
CASE
FLOC NOZZLE
SPLASH COMPARTMENT
POOL LEVEL
INTERNAL EFFLUENT SKIMMER
TRUNNION SEALS
SOLIDS DISCHARGE PORTS AND PLOWS
SOLIDS DISCHARGE
BASE
PILLOW BLOCK AND MAIN BEARING
EFFLUENT DISCHARGE
SKIMMER CONTROL
FLOC INLET
FEED INLET

FIGURE 7-19. Cross section of concurrent flow solid – bowl centrifuge.

Basket Centrifuge

The basket centrifuge or imperforate bowl-knife discharge unit, as shown in Figure 7-20, has recently been introduced primarily for use as a partial dewatering device at small plants. Parkhurst et al. [30] have used the basket centrifuge as a clarifying device for solid bowl centrate at the Los Angeles County Sanitary District.

Flow enters the machine at the bottom and is directed toward the outer wall of the basket. Cake continually builds up within the basket until the centrate, which overflows a weir at the top of the unit, begins to increase in solids. At that point, feed to the unit is shut off, the machine decelerates, and a skimmer enters the bowl to remove the liquid layer remaining in the unit. A knife is then moved into the bowl to cut out the cake which falls out the open bottom of the machine. The unit is a batch device with alternate charging of feed sludge and discharging of dewatered cake. Because of the cycle time involved this unit has a lower capacity than continuous devices. It does, however, have the capability of higher solids recovery without chemical addition because there is a minimum of disturbance of the depositing solids.

Disc Centrifuge

The disc centrifuge has long been used in the chemical process industry for handling large flows with relatively low concentrations of fine particles. The incoming stream is distributed between a multitude of narrow channels formed by stacked conical discs. Figure 7-21 shows a cross section of a disc centrifuge. Suspended particles have only a short distance to settle, so that small and low density particles are readily collected and discharge continuously through fairly small orifices in the bowl wall. Sludge concentrations of 5- to 20-fold are accomplished. The clarification capability and throughput range are high, but sludge concentration is limited by the necessity of discharging through orifices of 0.050 inches to 0.100 inches in diameter, which imposes an upper limit on the size of particle that can be handled by the disc centrifuge. Feed must be degritted and adequately screened. Insufficient attention to these factors in the past has led to the erroneous conclusion that disc centrifuges are not applicable to some industrial and municipal sludges due to plugging. Plugging is a real limitation, however, and must be considered in the design of a system employing disc centrifuges. Even if the sludge is screened adequately, severe nozzle clogging can occur if the feed to the centrifuge is stopped, interrupted, or reduced below some minimum value. The nozzles immediately clog due to the collapse of the solids built up in the bowl as described by Woodruff [31]. They are particularly of value in classifying the sludges resulting from lime coagulation of secondary effluent. For handling organic sludges, the thickening ability of the disc centrifuges is good but their dewatering ability leaves much to be desired.

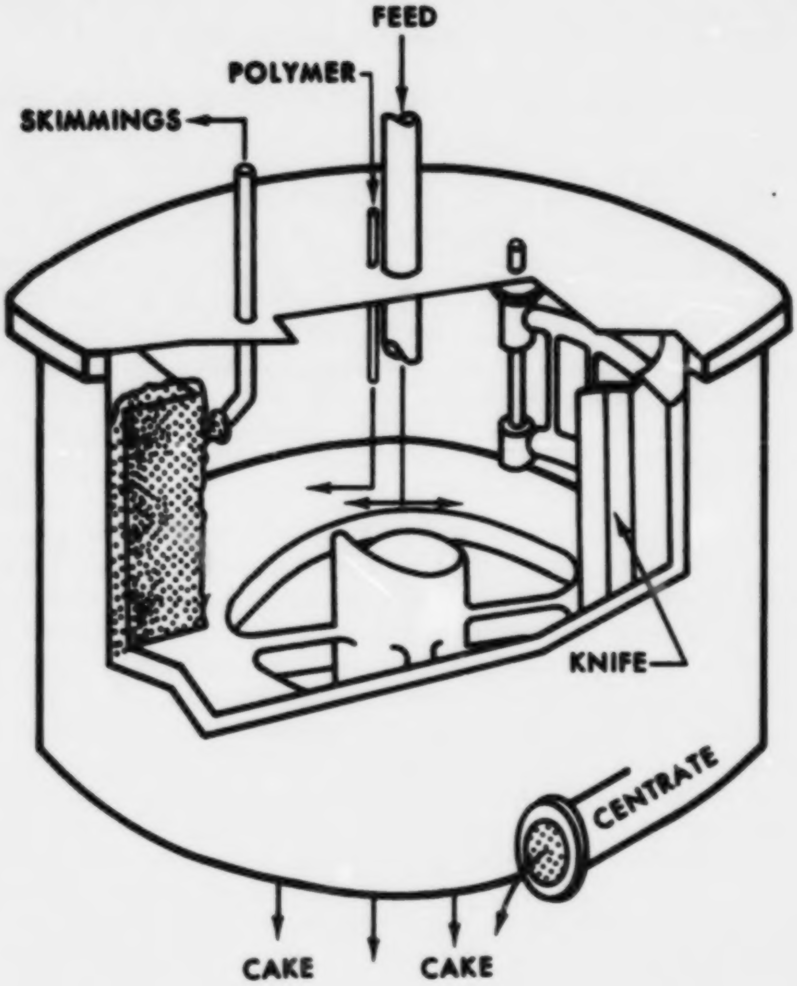

FEED
POLYMER
SKIMMINGS
KNIFE
CENTRATE
CAKE
CAKE

FIGURE 7-20. Schematic diagram of a basket centrifuge.

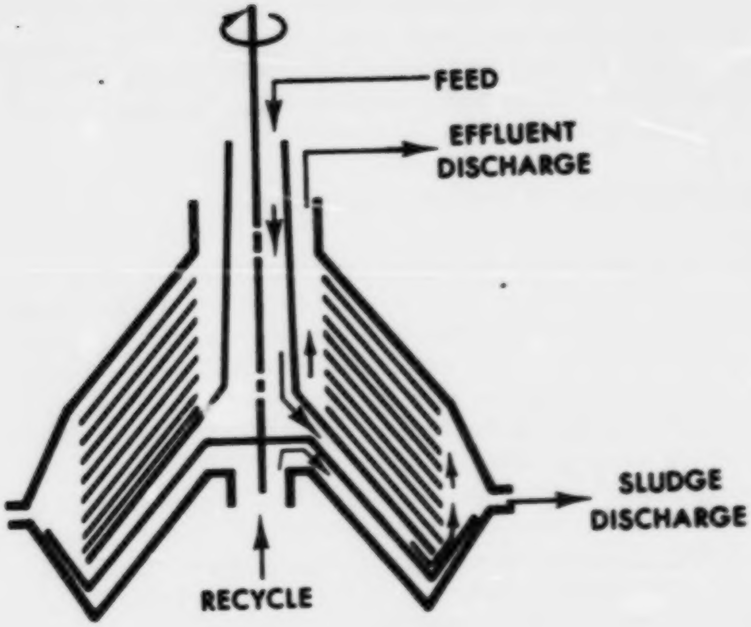

FEED
EFFLUENT DISCHARGE
SLUDGE DISCHARGE
RECYCLE

FIGURE 7-21. Disc type centrifuge.

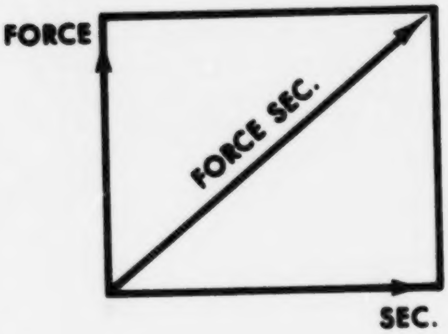
Lower Speed Continuous Solid Bowl Conveyor Type Centrifuges

These centrifuges have been developed primarily in Europe to achieve high solids capture and minimize the recirculation of solids to the treatment plant without the use of high polymer dosages. The sludge is introduced into the centrifuge with the lowest possible acceleration and turbulence. The machine is operated at about 1,500 rpm depending on the diameter of the centrifuge. This low rpm gives a low noise level and a minimum of wear and tear on the rotating parts. Low conveyor differential speeds are also used. Machines of this type are now offered by I. Krüger and Co. of Denmark and Bird Machine. Among the reported advantages of these machines are lower capital costs, lower power requirements, lower noise level, and reduced maintenance when compared to higher speed centrifuges. The use of large pool volumes, reduced internal turbulence and low centrifugal forces (500 to 800) combine to reduce shearing forces on the floc and improve the conveying characteristics.

7.3.3 Sludge Fractionation (Classification) by Centrifuge

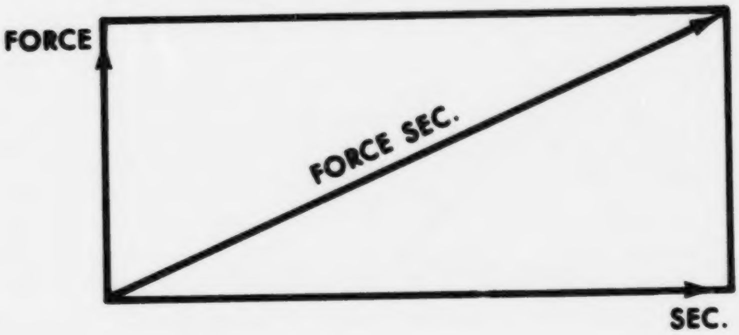
Excess activated sludges and hydrated fines tend to be lighter than primary sludge or well flocculated secondary sludges. The behavior of heavy and light sludge particles in centrifuges can be visualized from Figure 7-22. It is seen that light sludge particles require a much longer retention time. This characteristic has been used to fractionate sludge into a cake fraction containing essentially all the primary sludge particles and a centrate with the lighter, excess activated or fine hydrated particles. This phenomena was used as a means of venting the more difficult to process solids into the plant effluent where effluent standards permitted this practice. It is also used as a means of separating various types of sludges to permit disposal of the one stream and further processing of the other. The former use is exemplified by past practices at some coastal plants which utilize ocean disposal of effluent at points beyond the continental shelf. The latter use is exemplified by the centrifugal classification at South Lake Tahoe [32] and Contra Costa, California [33] plants for fractionation of lime sludges into a calcium carbonate-rich fraction and an inert laden fraction.

In addition to the primary effect of residence time, the setting of pool depth materially affects the degree of fractionation achieved. In tests at Los Angeles [30] on a digested primary sludge containing some sludge from satellite secondary plants, it was noted that the lower the solids capture, the drier the cake due to the venting of the fine hydrated particles in the centrate. The low level of solids recovery (less than or equal to 50 percent achieved with no chemical conditioning) is clearly inapplicable in most plants as indicated by the data in Figure 7-23. However, the work at Los Angeles was carried out to select the optimum system to cope with new effluent standards which could not be met by previous method and equipment. By using polymers (8 to 10 lb/ton) and a 3.4-inch pool depth, solids recovery of as high as 95 percent were achieved under the same operating conditions shown in Figure 7-23.



HEAVY SLUDGE PARTICLES

FORCE
FORCE SEC.
SEC.



LIGHT SLUDGE PARTICLES

FORCE
FORCE SEC.
SEC.

FIGURE 7-22. Reaction of particles within centrifuges.

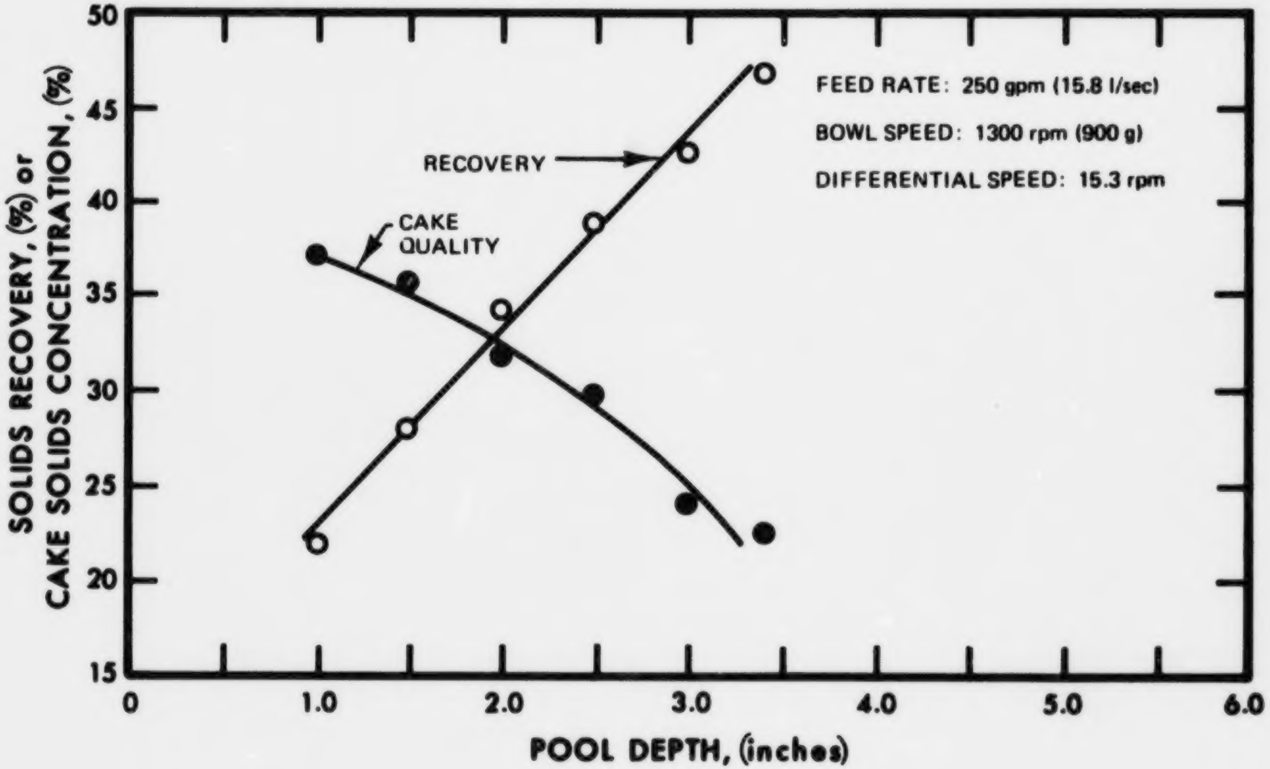


FIGURE 7-23. Typical dewatering performance curves for a 36 "X 96 " Bird horizontal scroll centrifuge fed unconditioned primary digested sludge [30].

At the South Tahoe [32] and Contra Costa [33] plants, solid bowl centrifuges are used on sludges resulting from lime coagulation of wastewaters. Phosphates and other inerts are removed with the centrate after first stage centrifugation while calcium carbonate is retained in the dewatered cake, which is later recalcined and reused. The first stage centrate is then dewatered in a second solid bowl centrifuge and the resultant cake is incinerated. By controlling the pool depth in the first stage centrifuge, it is possible to control the solids capture. If the first stage machine is operated to maximize solids capture (90 to 95 percent), then the cake contains nearly all of the inert materials as well as the calcium carbonate desired for the recalcining process. Figure 7-24 illustrates the classification effects on the percent recovery of various constituents of lime sludges resulting from coagulation of raw wastewaters at Contra Costa. On the tertiary lime sludges at South Tahoe, it was found that by operating the first centrifuge at 75 percent capture, the calcium oxide content of the cake was improved by 15 percent over that obtained at 95 percent. Such operation resulted in 87 percent of the usuable lime going to the recalcining furnace and the purging of 80+ percent of phosphorus and 35+ percent of magnesium from the recalcining system. Similar results were achieved at Contra Costa where 90 percent of the calcium carbonate fed to the first stage centrifuge was recovered while 50 to 75 percent of the other constituents were rejected in the centrate. The first stage cake had a solids content of 50 percent or greater.

7.3.4 System Requirements

Most centrifuge installations include the items shown in Figure 7-25. Sludge pumps are usually of the progressing cavity type, since constant feed is essential. Not shown, but frequently required where gritty sludge is encountered, is a cyclone separator for auxiliary degritting to prevent excessive wear. Details of a typical flocculant system are shown in Figure 7-26. Although not shown in this figure, it is preferable to have separate tanks for mixing of the flocculant and storage of the feed pump supply. This avoids feeding incompletely mixed flocculent solution. For solid bowl machines, the flocculant is usually added directly to the interior of the centrifuge to avoid shearing the floc. As discussed in detail in Chapter 6, the most effective flocculant and its dosage for any given sludge can only be determined by experimentation. Dilution of the flocculant ahead of the centrifuge to a strength of 0.1 percent or less has often been found to give maximum effectiveness. A dispersing eductor is normally used for this purpose.

7.3.5 Results of Centrifugal Dewatering

Data is readily available on the use of solid bowl centrifuges since they are widely used. As noted earlier, it is difficult to predict dewatering results. However, Table 7-6 presents typical data which can be expected with the solid bowl centrifugation of various organic sludges.

Figure 7-27 graphically depicts the effects of the major variables on solid bowl performance in dewatering an anaerobically digested mixture of primary and secondary

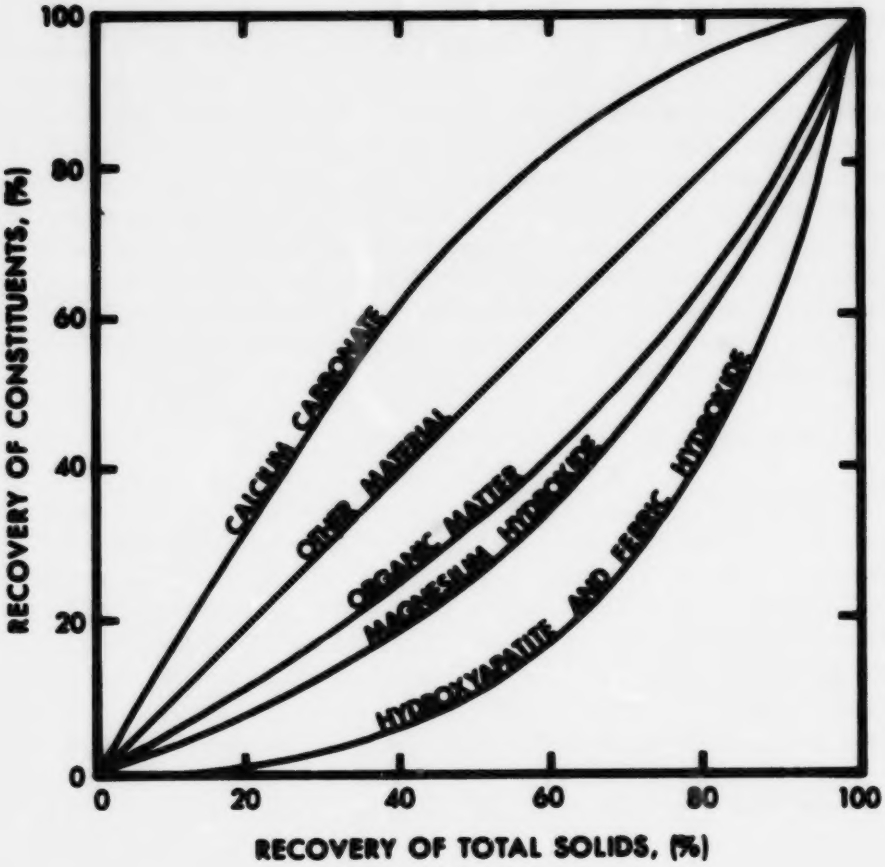


FIGURE 7-24. Summary of constituent recoveries during wet classification of lime sludges resulting from raw wastewater coagulation [33].

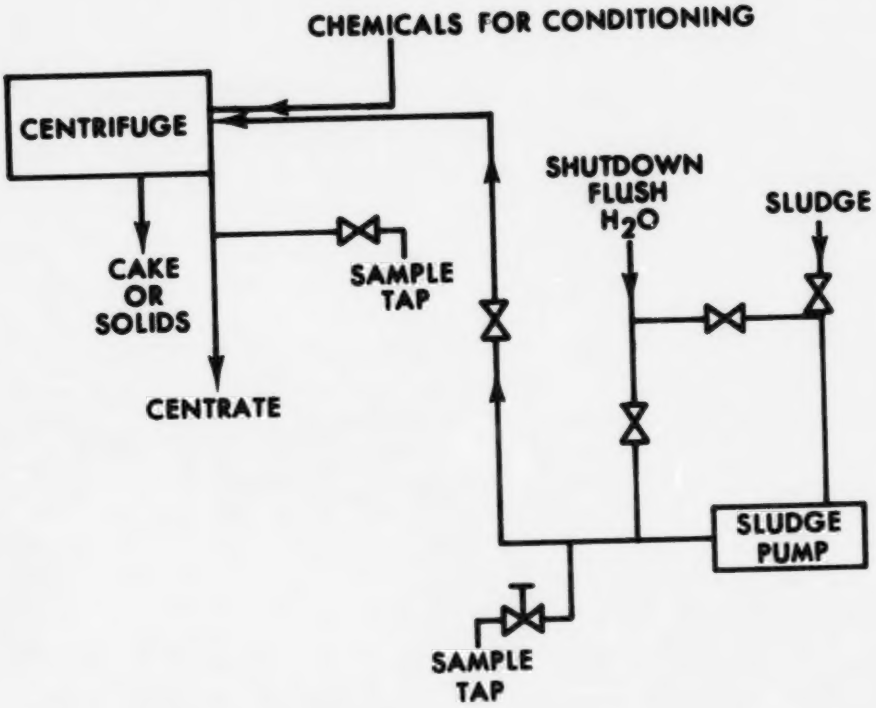

CHEMICALS FOR CONDITIONING
CENTRIFUGE
CAKE OR SOLIDS
CENTRATE
SAMPLE TAP
SHUTDOWN FLUSH H_2O
SLUDGE
SLUDGE PUMP
SAMPLE TAP

FIGURE 7-25. Centrifuge dewatering system.

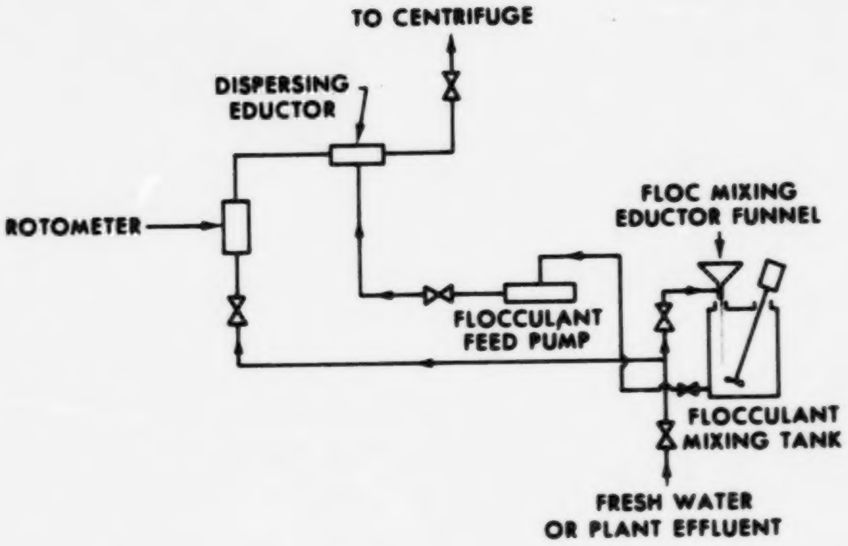

TO CENTRIFUGE
DISPERSING EDUCTOR
ROTOMETER
FLOC MIXING EDUCTOR FUNNEL
FLOCCULANT FEED PUMP
FLOCCULANT MIXING TANK
FRESH WATER OR PLANT EFFLUENT

FIGURE 7-26. Typical flocculant piping diagram.

TABLE 7-6

TYPICAL SOLID BOWL CENTRIFUGE PERFORMANCE [6]

	Sludge Cake Characteristics		
Wastewater Sludge Type	**Solids (%)**	**Solids Recovery (%)**	**Chemical Addition**
Raw or digested primary	28-35	70-90 (50-70)*	no
Raw or digested primary, plus trickling filter humus	20-30	80-95 60-75	yes no
Raw or digested primary, plus activated sludge	15-30	80-95 50-65	yes no

*New data indicate performance is in this range.

sludge. In this particular case the maximum percent solids recovery was well below 50 percent when the feed rate was 10 gpm and no flocculant was employed. However, the deeper pond depth with a feed rate of 10 gpm and chemical addition raised the solids recovery to 90 percent. The chemical cost for this improvement was about $5.50 per ton. Figure 7-28 shows the effect of centrifuge operation at various capture levels on cake solids concentration for the same case. At about 90 percent solids recovery, the cake solids varied from 17 to 18 percent for the particular sludge and polyelectrolyte used.

Data by Albertson and Sherwood [34] on digested primary sludges indicate that 96 percent solids recovery and a cake solids concentration of 28 percent are achievable, while at 82 percent solids recovery, a 36 percent cake solids concentration can be achieved. Townsend [35] concluded in his work that raw primary sludges dewater to 30 to 40 percent solids with the assistance of 1.5 to 2.5 lb/ton polymer, while digested primary dewaters to 20 to 30 percent solids with 3 to 6 lb/ton of polymer. Heat treated sludges have been dewatered to 35 to 45 percent solids and no polymer was required for 85 percent capture. Recoveries of 92 to 99 percent of the solids from heat treated primary sludges have been reported by Albertson [36] with polymer costs of $1.85 to $5.35 per ton of dry solids. Dewatering of heat treated mixtures of activated and raw primary sludge has produced cake solids concentrations of 40 percent with a 95 percent solids recovery. No chemicals were required. However, the use of $4.35 per ton of polymers in this latter case enabled a 50 percent increase in centrifuge capacity while producing cake solids of 28 percent.

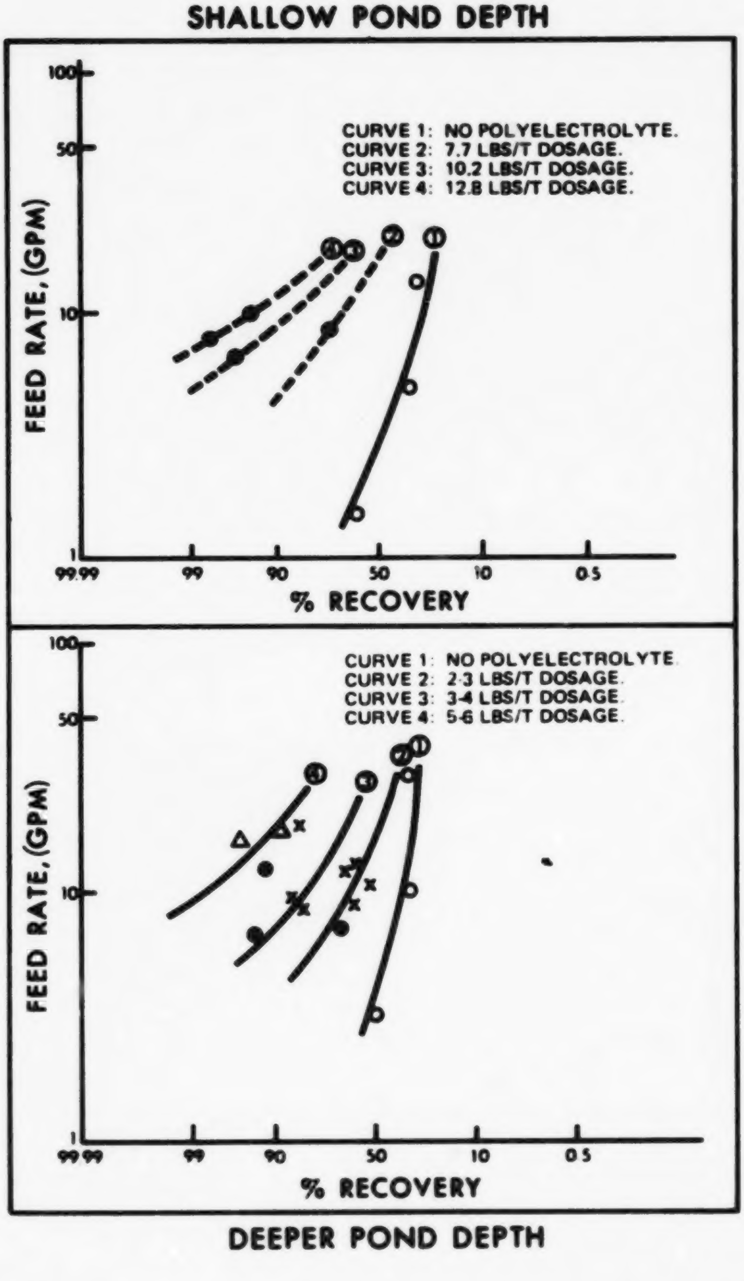
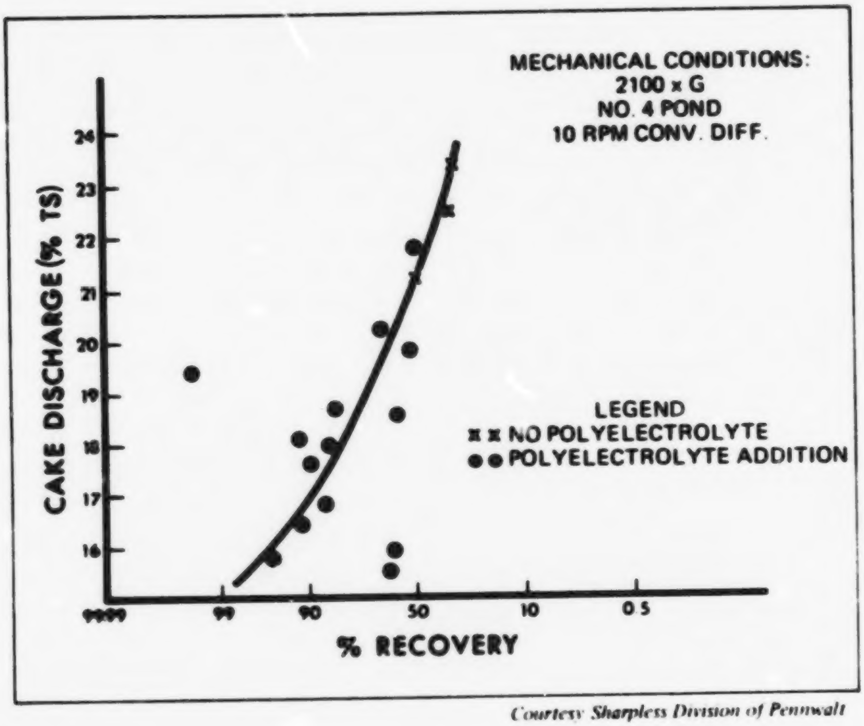


FIGURE 7-27. Effect of polyelectrolyte dosage and pool depth on percent solids recovery at various feed rates.

Courtesy Sharples Division of Pennwalt



Courtesy Sharpless Division of Pennwalt

FIGURE 7-28. Cake dryness as a function of solids recovery.

Table 7-7 presents data by Young, Matsch, and Wilcox [37] on the dewatering of air and oxygen EAS by solid bowl and basket centrifuges. These sludges are from systems treating raw wastewater or primary effluents. With both the scroll type and the basket type centrifuges, a slightly higher cake concentration may be achieved and less chemical conditioning may be required with oxygen excess activated sludge. Table 7-7 also illustrates that basket centrifuges are generally capable of providing drier activated sludge cakes than are the solid bowl machines because there is minimum disturbance of the depositing solids. Kyte [38] found that the basket centrifuge could provide a cake of 13 to 14.6 percent solids from a feed of 1 percent solids and capture 94.5 to 97.4 percent of the solids without polymer addition.

The basket centrifuges are capable of achieving a higher degree of dewatering than are disc centrifuges. A basket centrifuge will typically dewater an EAS from an initial solids concentration of 0.5 to 1.5 percent to a final solids concentration of 10 to 12 percent. With no chemical addition, a solids capture of 90 percent is possible. For the same sludge, a disc type centrifuge could only achieve a final solids concentration of 6 percent. While it is sometimes preferable to obtain a higher activated sludge concentration than the 5 to 6 percent produced by disc centrifuges, there are situations where this level of concentration is desirable. Excess activated sludge concentration of less than 1 percent unnecessarily reduces the volumetric capacity of an anaerobic digester by a factor of 6 or more. Similarly, wet oxidation or heat treatment processes, where expensive equipment is involved, can be more economically sized when fed at concentrations of at least 5 percent. The disc centrifuge has also been used by Woodruff [31] to thicken activated sludge prior to dewatering in a different centrifuge. By removing many of the fine particles in the disc machine, the solid bowl centrifuge can produce a 22 percent cake. A concentration of 13 to 19 percent was the best that could be achieved without prior thickening by the disc machine.

Another approach to series centrifugation for organic sludge dewatering was studied in Los Angeles [30] where objectionable amounts of floating and settleable material were present in the centrate from existing solid bowl machines. This material was discharged with the primary effluent to the ocean. The solid bowl machines operated without polymer on digested primary sludge and produced a cake of about 35 percent solids with 35 to 40 percent solids recovery. The solids recovery could be increased to 95 percent by the use of about 10 lb/ton of cationic polymer, but the cake solids concentration was reduced to 18 to 20 percent. However, variations in digested sludge characteristics resulted in erratic recoveries at even this polymer dosage. An alternate mode of operation studied was to pass the centrate from the solid bowl unit through a basket centrifuge and not use polymers. Solids recovery did not exceed 80 percent, which was inadequate. However, with a polymer dosage of as low as 2 lb/ton, 96 percent solids recovery was possible. The combined solid bowl and basket centrifuge cakes had a 25 percent solids concentration. In other tests by Keith and Little [39] it was found that the series approach with aerobically digested sludge and without polymer addition could achieve the same process performance as a single stage solid bowl machine operating with polymer doses of 7 to 13 lb/ton.

TABLE 7-7

DEWATERING OF OXYGEN ACTIVATED SLUDGES IN SOLID BOWL AND BASKET CENTRIFUGES [37]

	Feed Solids Concentration (% by wt)		Feed Rate (gpm)		Solids Capture (%)		Cake Concentration (% solids by wt.)		Polymer Addition (lb/ton dry solids)	
Centrifuge Type	Oxygen	Conventional	Oxygen	Conventional	Oxygen	Conventional	Oxygen	Conventional	Oxygen	Conventional
Solid Bowl										
Raw wastewater	2-3	1.5-2.5	50-60	45-55	85-90	80-85	10-13	9-11	3	6-10
Primary effluent	2-3	0.7-1.3	90-100	55-65	90-95	80-85	8-10	8-9	3	6-10
Basket										
Raw wastewater	2-3	1.5-2.5	35-40	20-35	92-96	90-95	9-12	9-11	0	0
Primary effluent	2-3	0.7-1.3	35-45	35-45	92-97	90-95	10-14	9-11	0	0

7.3.6 Summation

The successful adaption of centrifugal devices to the dewatering of sludges that contain a significant quantity of activated sludge is becoming common. Design improvements are increasingly aimed at obtaining a minimum of 90 percent solids capture with little chemical conditioning. The advent of the concurrent flow and the lower speed machines should materially aid the successful adaption of solid bowl centrifuges. Plant designs should be based upon scale-up of pilot tests whenever possible, and several manufacturers have pilot-scale centrifuges available for evaluation. The manufacturers will provide assistance in scaling up the pilot test data. In addition to dewatering functions, solid bowl, basket, or disc centrifuges can also provide classification of organic or chemical sludges.

7.4 Drying Beds

7.4.1 Factors Affecting Design

The most widely used dewatering method in the United States is drying of the sludge on open or covered sandbeds. Over 6,000 wastewater treatment plants use this method according to Burd [40]. Although they are especially popular in small plants, drying beds are also used by 38 percent of the cities serving populations of over 100,000. Furthermore, sandbed drying is the most common technique employed in Europe. Sandbeds possess the advantage of needing little operator attention and skill. Air drying is normally restricted to well digested sludge, because raw sludge is odorous, attracts insects, and does not dry satisfactorily when applied at reasonable depths. Oil and grease discharged with raw sludge clog sandbed pores and thereby seriously retard drainage. The design and use of drying beds are affected by many parameters. They include weather conditions, sludge characteristics, land values and proximity of residences, and use of sludge conditioning aids. Climatic conditions are most important. Factors such as the amount and rate of precipitation, percentage of sunshine, air temperature, relative humidity, and wind velocity determine the effectiveness of air drying.

The nature and moisture content of the sludge discharged to drying beds influences the drying process. Sludges containing grit dry rapidly; those containing grease more slowly; aged sludge dries slower than new sludge; primary sludge dries faster than secondary sludge; and digested sludge dries faster than fresh sludge and cracks earlier. It is important that wastewater sludge be well digested for optimum drying. In well digested sludge, entrained gases tend to float the sludge solids and leave a layer of relatively clear liquid, which can readily drain through the sand.

7.4.2 Design Criteria for Sandbeds

Design standards vary from region to region in the United States. This variance is partly due to climatic differences. However, Table 7-8 presents typical criteria for the design of open sand drying beds.

TABLE 7-8

CRITERIA FOR THE DESIGN OF SANDBEDS [15]

Type of Digested Sludge	Area (sq ft/capita)	Sludge Loading Dry Solids (lb/sq ft/yr)
Primary	1.0	27.5
Primary and standard trickling filter	1.6	22.0
Primary and activated	3.0	15.0
Chemically precipitated	2.0	22.0

Sandbeds can be enclosured by glass. Glass enclosures protect the drying sludge from rain, control odors and insects, reduce the drying periods during cold weather, and can improve the appearance of a waste treatment plant. Experience has shown that only 67 to 75 percent of area required for an open bed is needed for an enclosed bed. Good ventilation is important to control humidity and optimize the evaporation rate. As expected, evaporation occurs rapidly in warm, dry weather. Adaptation of mechanical sludge removal equipment to enclosured beds is more difficult than to open drying beds.

Drying beds usually consist of 4 to 9 inches of sand which is placed over 8 to 18 inches of graded gravel or stone. The sand typically has an effective size of 0.3 to 1.2 mm and a uniformity coefficient of less than 5.0. Gravel is normally graded from 1/8 to 1.0 inches. Drying beds have underdrains that are spaced from 8 to 20 feet apart. Underdrain piping is often vitrified clay laid with open joints, has a minimum diameter of 4 inches, and has a minimum slope of about 1 percent. Collected filtrate is usually returned to the treatment plant.

The Ten State Standards [41] make the following design recommendations.

- The top 3 inches of gravel consist of 1/8- to 1/4-inch gravel
- The gravel extend at least 6 inches above the top of the underdrains
- The gravel layer should be 12 inches deep
- The sand layer should be at least 6 to 9 inches deep

- Underdrains should be no more than 20 feet apart

- Bed walls should be watertight and extend 15 to 18 inches above and at least 6 inches below the surface

- Outer walls should be curved to prevent soil from washing onto the beds

- At least 2 beds should be provided

- Pairs of concrete truck tracks at 20-foot centers should be provided for all beds

- The influent pipe should terminate at least 12 inches above the surface with concrete splash plates provided at discharge points

Mechanical lifting of sludge has been practiced for many years at some large treatment plants, but now it is receiving more attention as the need to minimize problems with labor costs. Mechanical devices can remove sludges of 20 to 30 percent solids while cakes of 30 to 40 percent are generally required for hand removal.

7.4.3 Results of Sandbed Drying

Typical performance data show rapid dewatering during the first one to two days while water removal by drainage predominates. This is followed by two to five weeks of slow dewatering principally by evaporation. In good weather, a cake of 45 percent solids may be achieved in six weeks with a well digested sludge. As with any dewatering technique, the performance of a sandbed may be improved by proper chemical conditioning and the dewatering time may be reduced by 50 percent or more [40]. Liftable sludges have been achieved in less than one day with proper chemical conditioning. Solids contents of 85 to 90 percent have been achieved on sandbeds.

7.4.4 Other Types of Drying Beds

In recent years, there has been increased interest in the use of paved drying beds with limited drainage systems for dewatering wastewater sludges. The original reasons for proposing paved bed design was to permit the use of mechanical equipment for cleaning the beds and thereby reduce the cost of labor and sand replacement. The feasibility of this approach has been demonstrated by field experience reported by South [42]. This experience indicates that modification of drying beds to provide only limited drainage facilities does not necessarily impare bed performance. In fact, the use of paved drying beds with limited drainage resulted in shorter drying times as well as more economical operation when compared with conventional sandbeds. The shorter time occurred because the use of mechanical equipment for cleaning permits the removal of sludge with a higher moisture content than in the case of hand cleaning. Paved beds have worked successfully with anaerobically digested sludges. Randall [43] and Randall and Koch [44] determined that the use of paved beds of center drain design for dewatering aerobically digested activated sludge is not as desirable as conventional sandbeds. Lateral drainage of activated sludge on a paved bed is very poor and does not contribute significantly to the dewatering process. The supernatant overlying the sludge cake may drain but the reduced drainage area of the paved bed compared to the sandbed greatly hampers the rate of drainage for aerobic sludges.

Wedge wire drying beds have been used sucessfully in England as reported by Burd [40] and Gauntlett and Packham [45] and a cross-sectional view of one is shown in Figure 7-29 by Crockford and Sparham [46]. This approach prevents the rising of water by capillary action through the media and the construction lends itself well to mechanical cleaning. The first United States installations have been made at Rollinsford, New Hampshire, and in Florida. The procedure used for dewatering sludge begins with the placement of water or effluent to a depth of up to 1 inch over the wedge wire. This water serves as a cushion permitting the added sludge to float without causing upward or downward pressure across the wedge wire surface and prevents compression or other disturbance of colloidal particles. The water is then run off by opening the control valve and controlling the rate so as to prevent a turbid effluent from the bed. After the free water has been drained, the bed is allowed to dry by drainage and evaporation until the sludge can be removed. It is possible, in small plants, to place the entire dewatering bed in a tiltable unit from which sludge may be removed merely by tilting the entire unit mechanically. Some typical performance data from the application of chemically conditioned sludges to wedge wire units are shown in Table 7-9.

TABLE 7-9

TYPICAL PERFORMANCE DATA [46]

Sludge Type	Feed Solids (%)	Cake Solids (%)	Time Interval	Solids Capture (%)
Primary	8.5	25	14 days	99
Trickling Filter Humus	2.9	8.8	20 hours	85
Digested Primary + EAS	3.0	10.0	12 days	86
Fresh EAS	0.7	6.2	12 hours	94
Fresh EAS	1.1	9.9	8 days	87
Thickened EAS	2.5	8.1	41 hours	100

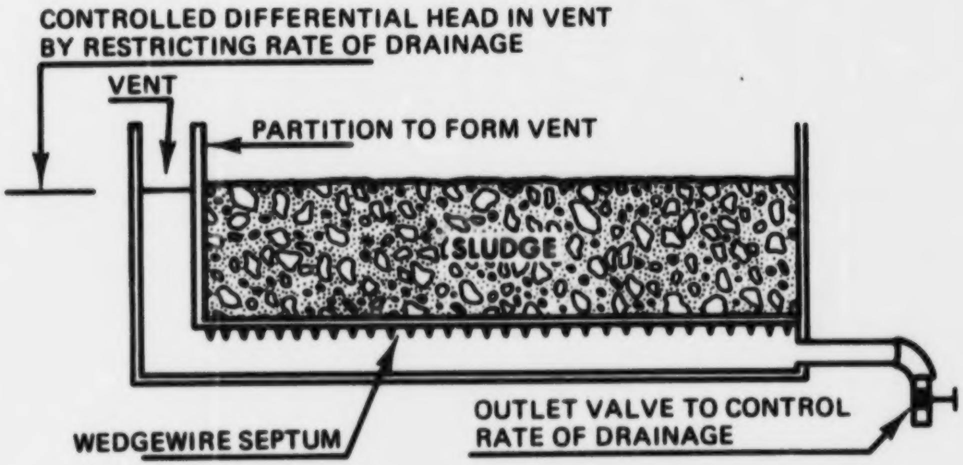

CONTROLLED DIFFERENTIAL HEAD IN VENT
BY RESTRICTING RATE OF DRAINAGE
VENT
PARTITION TO FORM VENT
SLUDGE
WEDGEWIRE SEPTUM
OUTLET VALVE TO CONTROL
RATE OF DRAINAGE

FIGURE 7-29. Cross section of a wedgewire drying bed [46].

The Rollingford, New Hampshire, plant reports that it dewaters excess activated sludge conditioned with polymers from two percent solids to a liftable condition in four hours.

7.5 Drying Lagoons

7.5.1 Factors Affecting Design

Lagoon drying is a low cost, simple system for sludge dewatering that has been commonly used in the United States. Drying lagoons are similar to sandbeds in that the sludge is periodically removed and the lagoon refilled. Lagoons have seldom been used where the sludge is never removed, because such systems are limited in application to areas where large quantities of cheap land are available. Sludge is stabilized to reduce odor problems prior to dewatering in a drying lagoon. Odor problems with lagoons can be greater than with sandbeds, because sludge in a lagoon retains more water for a longer period than does sludge on a conventional sand drying bed.

Other factors affecting design include consideration of groundwater protection and access control. Major design factors include climate, subsoil permeability, lagoon depth, loading rates, and sludge characteristics. The design should provide for uniform distribution of the sludge and for decanting of supernatant to speed the drying process.

7.5.2 Design Criteria for Drying Lagoons

Criteria cited in the Ten State Standards [41] are given below:

- The soil must be reasonably porous, and the bottom of the lagoon must be at least 18 inches above the maximum groundwater table.
- Surrounding areas must be graded to prevent surface water from entering the lagoon.
- The lagoon depth should not be more than 24 inches.
- At least two lagoons should be provided.

Underdrains have often been proposed where permeable soils are unavailable. However, evaporation is the main mechanism for water removal from a lagoon.

Solids loading rates suggested for drying lagoons are 2.2 to 2.4 lb/yr/cu ft of lagoon capacity [41]. Other designers have made recommendations ranging from 1 sq ft/capita for primary digested sludges in an arid climate to as high as 3 to 4 sq ft/capita for activated sludge plants where the annual rainfall is 36 inches. A dike height of about 2 feet with the depth of sludge after decanting of 15 inches has been used [6]. Sludge depths of 2.5 to 4 feet may be used in warmer climates where longer drying periods are possible [40]. Dikes should be of a shape and size to permit maintenance, mowing, and trucks and front-end loaders to enter the lagoons for sludge removal.

7.5.3 Results of Lagoon Drying

Sludge will generally not dewater in any reasonable period of time to the point that it can be lifted by a fork except in an extremely hot, arid climate. If sludge is placed in depths of 15 inches or less, it may be removed with a front-end loader in 3 to 5 months. When sludge is to be used for soil conditioning, it may be desirable to stockpile it for added drying before use. One proposed approach [40] utilizes a 3-year cycle in which the lagoon is loaded for 1 year, dries for 18 months, is cleaned, and allowed to rest for 6 months. Definitive data on lagoon drying are scarce. Sludge may be dewatered from 5 percent solids to 40 to 45 percent solids in 2 to 3 years using sludge depths of 2 to 4 feet [40]. In England, dewatering beyond 30 percent solids in lagoons is rare.

7.6 Pressure Filtration

7.6.1 Concept

The plate and frame filter press is a batch device, which has been used in industry and in European wastewater plants for many years to process difficult to dewater sludges. The press consists of vertical plates which are held rigidly in a frame and which are pressed together between a fixed and moving end as illustrated in Figure 7-30. On the face of each individual plate is mounted a filter cloth as shown in Figure 7-31 [47]. The sludge is fed into the press and passes through feed holes in the trays along the length of the press. The water passes through the cloth, while the solids are retained and form a cake on the surface of the cloth as explained by Brossman and Jensen [48]. Sludge feeding is stopped when the cavities or chambers between the trays are completely filled. Drainage ports are provided at the bottom of each press chamber. The filtrate is collected in these, taken to the end of the press, and discharged to a common drain. At the commencement of a processing cycle, the drainage from a large press can be in the order of 2,000 to 3,000 gallons per hour. This rate falls rapidly to about 500 gallons per hour as the cake begins formation and when the cake completely fills the chamber, the rate is virtually nothing. The dewatering step is completed when the filtrate is near zero. At this point the pump feeding sludge to the press is stopped and any back pressure in the piping is released through a bypass valve. The electrical closing gear is then operated to open the press. The individual plates are next moved in turn over the gap between the plates and the moving end. This allows the filter cakes to fall out. The plate moving step can be either manual or automatic. When all the plates have been moved and the cakes released, the complete pack of plates is then pushed back by the moving end and closed by the electrical closing gear. The valve to the press is then opened, the sludge feed pump started, and the next dewatering cycle commences.

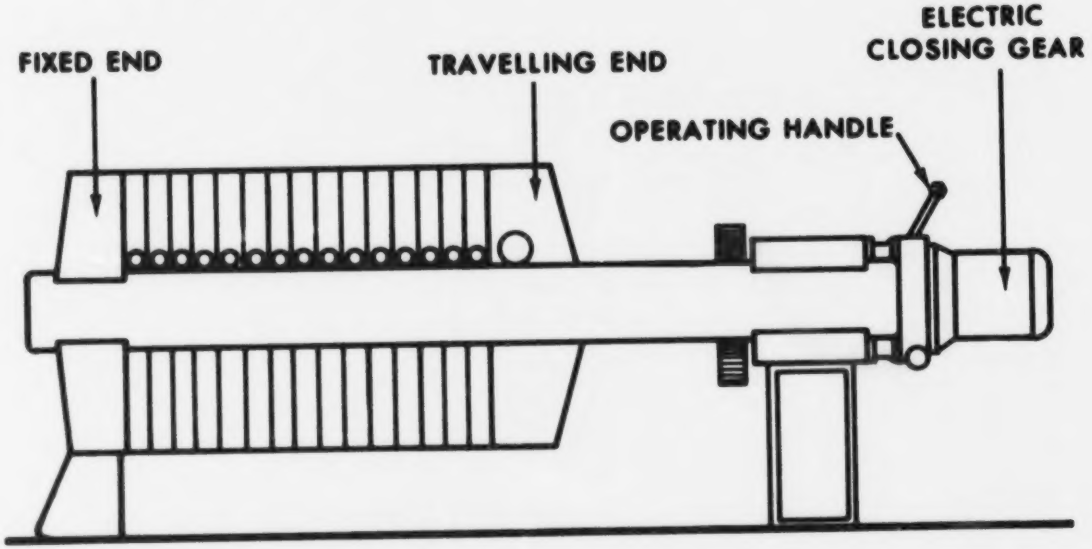

FIXED END
TRAVELLING END
ELECTRIC
CLOSING GEAR
OPERATING HANDLE

FIGURE 7-30. Side view of a filter press [47].

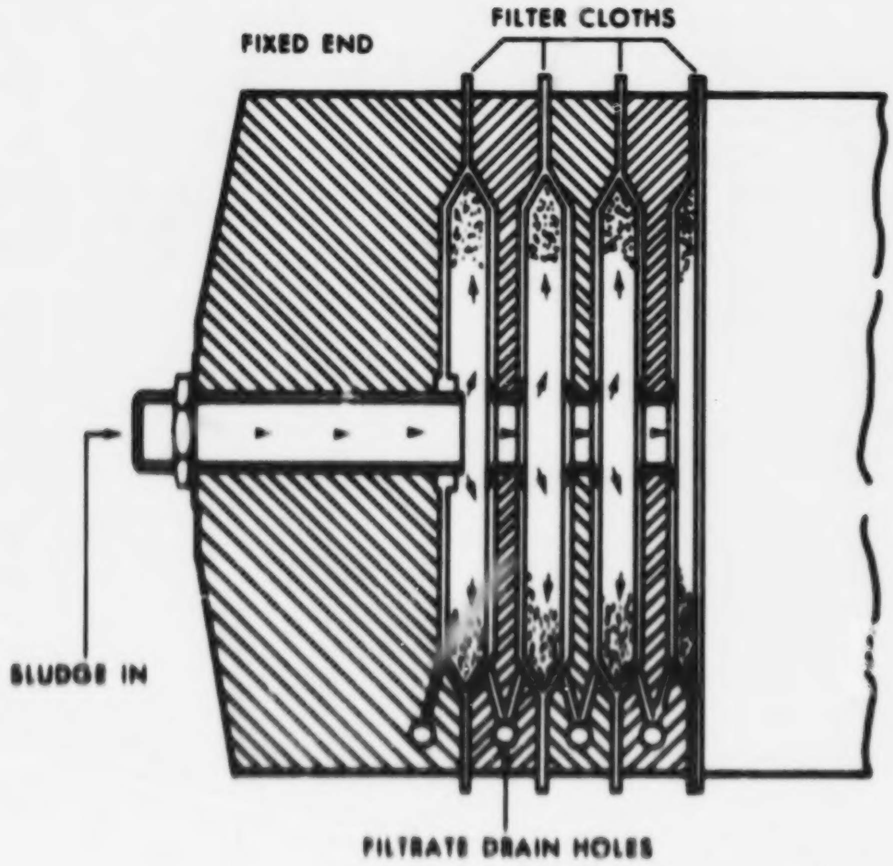

FIXED END
FILTER CLOTHS
SLUDGE IN
FILTRATE DRAIN HOLES

FIGURE 7.11. Cutaway view of a filter press [47].

Filter presses are normally installed well above floor level, so that the cakes can drop onto conveyors or trailers positioned underneath the press. The dry cake release may be assisted by introducing compressed air behind the filter cloth on both sides of the plate. This causes the cloth to flex and dislodge the cake. Although such a technique removes most of the cake, observations of operating presses indicate some manual attention may still be required to remove all of the cake. The pressures which may be applied to a sludge for removal of water by the filter presses now available range from 5,000 to 20,000 times the force of gravity. In comparison, a solid bowl centrifuge provides forces of 3,500 g and a vacuum filter, 1,000 g. As a result of these greater pressures, filter presses may offer several advantages as explained by Thomas [49]. Advantages and disadvantages of pressure filtration are listed in Table 7-10.

TABLE 7-10

PRESSURE FILTRATION CONSIDERATIONS

Advantages	Disadvantages
Higher cake solids concentrations (30 to 50 percent)	Batch operation
Improved filtrate clarity	High labor costs
Improved solids capture	Filter cloth life limitations
Reduced chemical consumption	Operator incompatibility
	Cake delumping

The reasons for past limited acceptance in the United States municipal field have been associated by Forster [50] with filter media, capacity, and discharge. At one time, the only filter medium available was a type of canvas duck. Such cloth would blind very rapidly with slimy sludges and had to be washed frequently. This manual operation added appreciable cost to the operation. The maximum size of filter plates available were 36" X 36", requiring an unusually long filter. In most cases, multiple installations were required to provide sufficient capacity for processing the quantities of sludge produced in a wastewater treatment plant. The opening and closing as well as the discharge of the accumulated filter cake was manual and required at least two men per filter press. This labor cost, in addition to frequent filter cloth washing, made the use of the machine unattractive. After World War II, filter press manufacturers reviewed the existing designs with a view to the elimination of these shortcomings. Manual closing and opening of filter presses is now accomplished by hydraulic cylinders and pumping systems. Mechanized plate shifting devices capable of moving a filter plate every three to four seconds has replaced manual plate shifting and has made it possible for one operator to care for several presses. The filter plate itself has received critical review. Leakage during operation was objectionable and the relatively thick cotton duck filter medium, which doubled as a gasket, was unsatisfactory. Present day filter plates are furnished with proper gaskets installed in machined grooves and permit leak-proof operation of such filters.

An increase in the acceptance of pressure filters is due to the introduction of suitable monofilament filter media. These materials, unlike multifilament filter cloth, do not readily blind in service. The latter is woven from a yarn and will slightly increase in diameter during service, because ultrafine solids are forced into its voids. The resultant decrease in the size of the cloth openings reduces the flow of filtrate through the media. No amount of washing will restore multifilament cloth to its original quality and porosity. The use of monofilament media has virtually eliminated this problem and greatly extended the filter cloth life. Many systems utilize an efficient precoat system for the filter. Such a system allows for the deposition of a protective layer of porous material on the filter media, prior to the start of the dewatering step. This prevents premature blinding. In sludge dewatering, ash from an incinerator, fly ash from a fossil fuel power plant, cement kiln dust, as well as buffing dust from a tannery, have been used successfully as precoat materials. In addition to protecting the filter cloth from blinding, the precoat acts as a parting plane and assures complete cake release during discharge. However, it is necessary to wash even the monofilament media periodically. This task was made easy and efficient by the development of mechanized plate washing devices.

Earlier filter presses had relatively small plate sizes (about 26 sq ft). Filter plate size was limited because the plate had to be shifted manually for cake discharge. However, with the advent of completely mechanized plate shifting, a much larger plate can be built and filtration capacity increased. This makes it possible to economically apply pressure filters in large sludge dewatering projects. While pressure filters with a total effective filtration area of 2,500 sq ft were once considered large, today units with an effective filtration area of 4,500 sq ft are not uncommon.

Until recently, pressure filters, with few exceptions, operated at a maximum pressure differential of 100 psi. Extensive studies during the early 1960s showed that pressure differentials of up to 225 psi produced filter cake solids concentration well in excess of 50 percent. Some commercially available systems now operate near these pressures.

7.6.2 System Requirements

Sludge is normally conditioned by the addition of inorganic chemicals and, sometimes, ash before pressure filtration. A dewatered cake shredder or delumper is needed when the cake will be further processed by a process like incineration [51]. Figure 7-32 shows the components of a typical system.

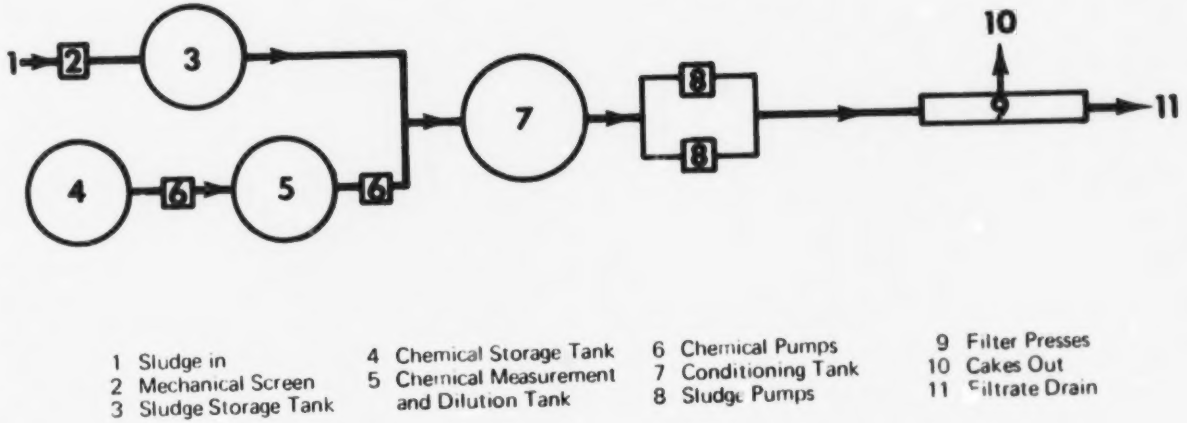

1
2
3
4
5
6
6
7
8
8
9
10
11
1 Sludge in
2 Mechanical Screen
3 Sludge Storage Tank
4 Chemical Storage Tank
5 Chemical Measurement and Dilution Tank
6 Chemical Pumps
7 Conditioning Tank
8 Sludge Pumps
9 Filter Presses
10 Cakes Out
11 Filtrate Drain

FIGURE 7-32. Filter press system [47]

7.6.3 Results of Pressure Filtration

Experience in the United States with pressure filtration of municipal sludges has been limited. Table 7-11 summarizes experiences on the wide variety of sludges treated by this technique in Europe. It is apparent that a 50 percent cake solids concentration can readily be produced with any sludge. The three variables of importance shown here are the quantity and type of chemical conditioner needed and cycle time. Of course ash is free, and the quantities of inorganic chemicals required seem to vary greatly from sludge to sludge. Two of the most significant United States filter press installations have recently been put into operation. One dewaters alum sludges from the treatment of water in Atlanta, Georgia [52,53], while the other dewaters a municipal wastewater sludge at Cedar Rapids, Iowa [54]. Other major installations are now in the design stage. They include facilities for dewatering tertiary lime and primary-secondary organic sludges at the Tahoe-Truckee Sanitation Agency's 6 mgd plant in California and for dewatering organic sludges and the rejects from a tertiary lime classifying centrifuge for a 60 mgd Washington Suburban Sanitary Commission facility in Montgomery County, Maryland.

Extensive pilot dewatering tests were conducted at Atlanta, Georgia, with alum sludge [illegible]. The data from these tests served as the design basis for the full-scale filter press installation. This facility can process 55,000 pounds/day dry solids. Filtration of the alum sludge without conditioning agents was impractical. Fly ash was an excellent conditioner. At a fly ash to alum ratio of 1:1 a cake of 60 percent solids was achieved in a 165 minute filtration cycle time and a pressure of 225 psi. Fly ash was also used as a precoat. Lime also provided excellent conditioning results. Quicklime addition in the amount of 13.5 percent by weight of the dry sludge solids produced a cake with 39.5 percent solids in a cycle time of 165 minutes. When hydrated lime added in the range of 10 to 15 percent of the sludge solids, filter cakes of 26 to 31 percent resulted in 75 to 105 minutes. The use of anionic polymers in amounts up to an equivalent cost of $3/ton of dry solids did not produce good cakes. Pulverized clay was also tested and found to be ineffective because it formed a thin, blinding film on the filter medium.

The initial operating results from the full-scale installation at Atlanta are consistent with the above pilot results. Cakes of 40 to 50 percent solids have been produced using a 10 percent lime dosage and precoating. The filters are 44-chamber machines with a capacity of 110 cu ft. They operate at a pressure of 225 psig and have a total area of 1,813 sq ft. The design cycle is 90 minutes which provides a gross loading of 3.16 lb/sq ft/cycle at an anticipated solids of 5 percent. A diatomaceous precoat of 100 lb/cycle is used. Typical filtrate suspended solids of less than 10 mg/l are reported. The March, 1969, bid price for the filter press plant was $2,807,560.

Cedar Rapids, Iowa, carried out an extensive pilot study on dewatering of an anaerobically digested mixed sludge from a 28.6 mgd two-stage trickling filter plant. Data from this study were used in the design of a sludge processing system which included conditioning, pressure filtration, and incineration. This system is now installed and uses a

TABLE 7-11

TYPICAL FILTER PRESS PRODUCTION DATA [50]

Sludge Type	Suspended Solids (%)	Conditioning of Dry Solids (%)	Cake Solids (%)	Time Cycle (hr)
Raw Primary	5-10	Ash 100	50	1.5
		FeCl$_3$ 5 Lime 10	45	2.0
Raw Primary with less than 50% EAS	3-6	Ash 150	50	2.0
		FeCl$_3$ 5 Lime 10	45	2.5
Raw Primary with more than 50% EAS	1-4	Ash 200	50	2.0
		FeCl$_3$ 6 Lime 12	45	2.5
Digested and Digested with less than 50% EAS	6-10	Ash 100	50	1.5
		FeCl$_3$ 5 Lime 10	45	2.0
Digested with more than 50% EAS	2-6	Ash 200	50	1.5
		FeCl$_3$ 7.5 Lime 15	45	2.5
EAS	Up to 5	Ash 250	50	2.0
		FeCl$_3$ 7.5 Lime 15	45	2.5

combination of recycled ash and chemical conditioning for dewatering of 56,000 lb of digested sludge per day [54]. The two pressure filters at Cedar Rapids are Beloit Passavant units and each has 83 chambers. The chambers have a volume of approximately 2.5 cu ft each. The filters each have 3,400 sq ft of filter area and are equipped with an expanded frame. This frame can hold as many as 100 chambers, which allows for further increases in sludge volume. The filter plates are carbon steel construction and epoxy coated, while the media is a monofilament polypropylene with a stainless steel mesh backup screen and a carbon steel drainage screen. The filter operates at approximately 225 psig pressure, and the plates, during operation, are held closed by hydraulic cylinders operating with an oil pressure at about 5,000 psig. The pressure filter plates are gasketed with asbestos Teflon gasket material, and the filter media is caulked in place with cotton caulking cord. Data reported on the Cedar Rapids mixed sludge indicate that it is very difficult to dewater because of a high and variable industrial loading on the plant. The ash/sludge solids ratio required for effective dewatering at Cedar Rapids is given in Figure 7-33. Thus, in order to function at an available and sustainable feed sludge solids level of about 4 to 5.25 percent, from 1.70 to 2.25 pounds of ash per pound of organic sludge was required. Further, the chemical requirements also varied with the sludge feed solid content, and this is as shown in Figure 7-34. It is obvious from both Figures 7-33 and 7-34 that the most economical operation occurs with a concentrated feed sludge. When the solids concentration decreases below about 5 percent, chemical requirements rapidly escalate.

The pressure filter at Cedar Rapids has proven to be a very efficient solids liquid separator. During the months of February to August, 1972, the filtrate averaged 74 mg/l of suspended solids, while the sludge feed solids averaged 4.6 percent, and the percent removal of suspended solids was essentially 100 percent. The filtrate has had an average COD of 857 mg/l, a BOD of 504 mg/l, and a total phosphorus concentration of less than 20 mg/l. Sludge cake discharges from the filter and is about 58 inches in diameter, 1½ inches thick, and weighs about 200 pounds. Each pressure filter produces 83 cakes per cycle, and these are discharged to a storage bunker. Shear cables are positioned across the top of the bunker to break the large cakes into smaller pieces. Cake is conveyed through a series of drag conveyors and elevators to the incinerator. At normal operation cake moisture content is in the range of 36 to 38 percent, and the cake's appearance is dense, dry, and textured.

Although the Cedar Rapids plant normally dewaters digested sludge, full-scale tests have been run on raw primary sludge. The same chemical dosages were required as for digested sludge with an ash to sludge solids ratio of 1 to 1. Filter cakes of 54 to 58 percent solids were achieved with filtrate solids being comparable to those achieved with digested sludge. However, the filtrate COD averaged 7,000 mg/l and the BOD averaged 5,700 mg/l. These values are much higher than those experienced with the dewatering of digested sludge.

The costs at Cedar Rapids are related to the concentration of feed solids. The average total capital and operating costs ranged from $26.83 per ton at 4.5 percent solids to $18.20 per ton at 6.5 percent solids. A breakdown of costs at 5.5 percent solids is as follows:

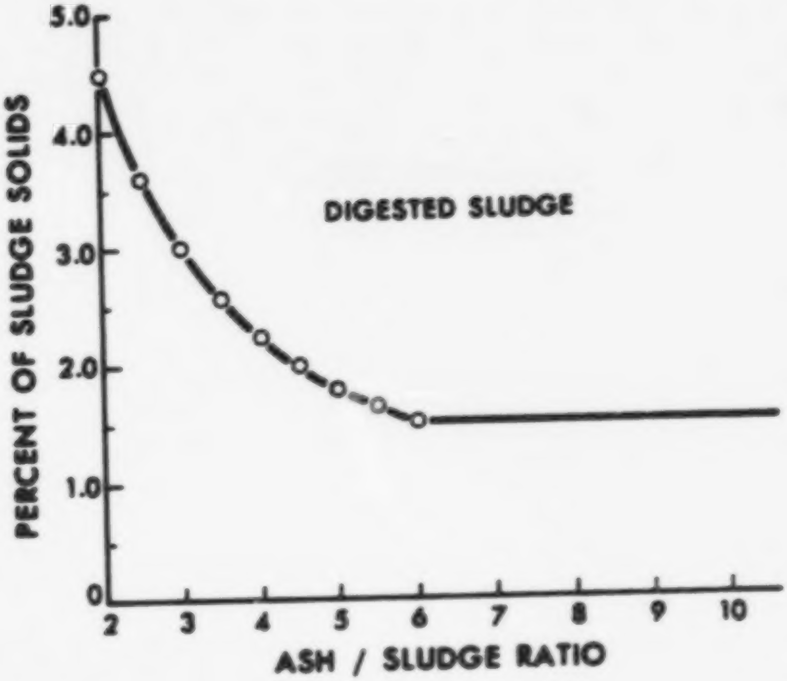


FIGURE 7-33. Required ash to sludge ratio as function of feed solids [54].

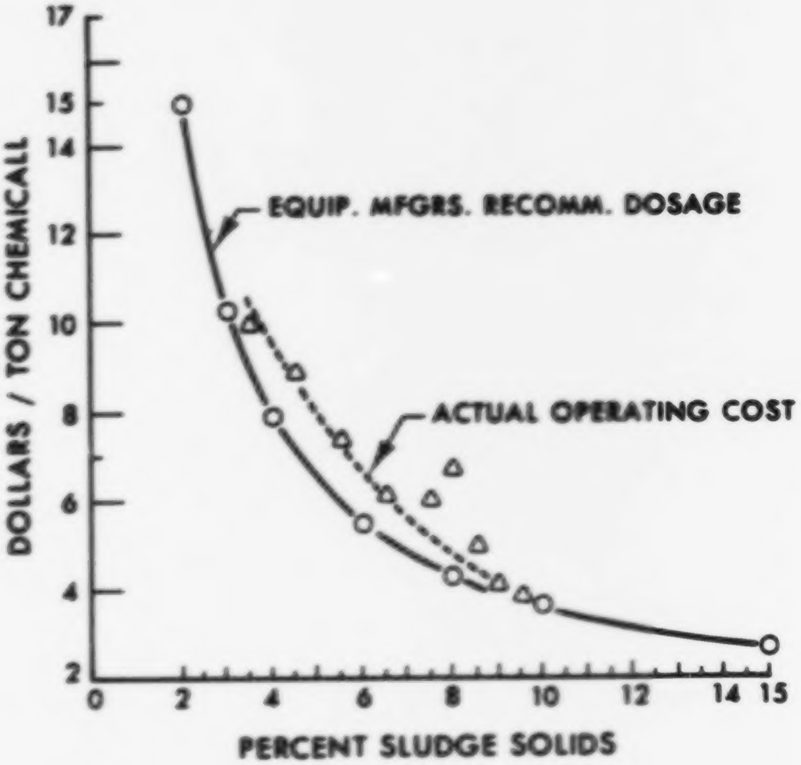


FIGURE 7-34. Average chemical costs for pressure filtration at Cedar Rapids [54].

Factor	**Average Dewatering Cost $/ton Dewatered**
Operating	4.69
Capital	9.71
Chemical	7.29
Total	21.69

Incineration has normally been achieved without the use of supplemental fuel. The filter has operated for 150 to 200 hours between media washings. Washing is achieved with a high pressure (750 psi) nozzle and a cleaning solution containing commercial grade detergent. Some operating difficulties have been experienced at Cedar Rapids, and pressure filter plate warpage has been a major problem. Warpage occurring in the plate diaphragm transfers bending to the plate frame, which in turn accelerates plate gasket deterioration due to a warped gasket seating plane. Plate alignment also affects gasket life, wherein poor alignment, or repositioning of the plates each cycle, causes a reshaping of the plate gasket leading to premature failure.

7.6.4 Summation

Recent improvements in the degree of automation, filter media, and unit capacities have led to renewed interest in pressure filtration for application to municipal sludges. Two major systems are in operation and several others are in the design stage. The ability to produce a very dry cake and clear filtrate are major points in favor of pressure filtration. The cake removal cycle still requires manual assistance.

7.7 Other Systems

Several types of dewatering devices which do not fall into the categories previously discussed are available and in use. These devices include:

- Moving screen concentrators
- Belt pressure filters
- Capillary dewatering systems
- Rotating gravity concentrators

7.7.1 Moving Screen Concentrator

This device is marketed by Smith and Loveless Co. [55]. Thickened and polymer treated sludge is distributed on a two-stage variable speed moving screen. Gravity dewatering occurs on a first moving screen, and the partially dewatered sludge then moves onto a second moving screen. It is passed through multiple compression stages for final dewatering. The compression dewatering is accomplished by passing the solids through sets of compression rollers with each successive set applying a higher pressure to the solids. Figure 7-35 shows the moving screen concentrator. The concentrator system depends on thickening of sludges by gravity using polymer addition, when straight activated sludge is being processed. The conditioner may not be required in the thickening step for other sludges.

The primary stage of the two-stage dewatering unit consists of an endless horizontal belt filter screen driven by a variable speed gear motor. This filter screen is made from a relatively open mesh, monofilament polyester material. The primary screen travels around two end rolls, one of which serves as the drive roll. A third roll is located midway between the other two rolls at a slightly elevated position. This flexing roll serves a dual purpose. It flexes the sludge cake to allow additional drainage of free water and it forms a conditioned sludge in the first section of the primary stage. The area where the sludge is introducted onto the first screen is designed to create a pool. In the pool, adequate head of one to two inches of water exists so that a thick cake is formed. Most free water drains through the screen in this area. As the screen travels on over the flexing roll, additional drainage of free water occurs. A drainage pan beneath the screen collects all the filtrate water and empties it into a central waste line. Finally, the concentrated sludge passes down a transition chute to the second dewatering stage. The primary screen is washed on its return trip to the sludge introduction area, so that a clean surface is always exposed to the incoming conditioned sludge. The open mesh filter screen facilitates complete washing.

The secondary stage of the dewatering unit consists of another horizontal, endless filter screen driven by a second variable speed gear motor. This stage is completely independent of the primary unit. The secondary filter screen is identical to that of the primary unit. This screen travels around two end rolls and between three sets of compression rollers. Each set of compression rollers can be adjusted to apply the desired pressure to the sludge being dewatered. The amount of pressure applied to the sludge depends upon the internal strength of the sludge and is generally highest for primary sludges and lowest for digested biological sludges. Also, the drier the sludge, the greater its strength and, for this reason, more pressure is applied with each successive set of pressure rollers. The sludge loses more free water with each compression, until it passes out from the last set of compression rolls, where relatively little free water remains in the sludge cake. This sludge cake then passes out of the secondary unit, down the discharge chute and into a dump truck or other receptacle. The results obtained using a 40-inch wide screen at several different locations and with various sludge types are summarized in Figure 7-36. Required polymer

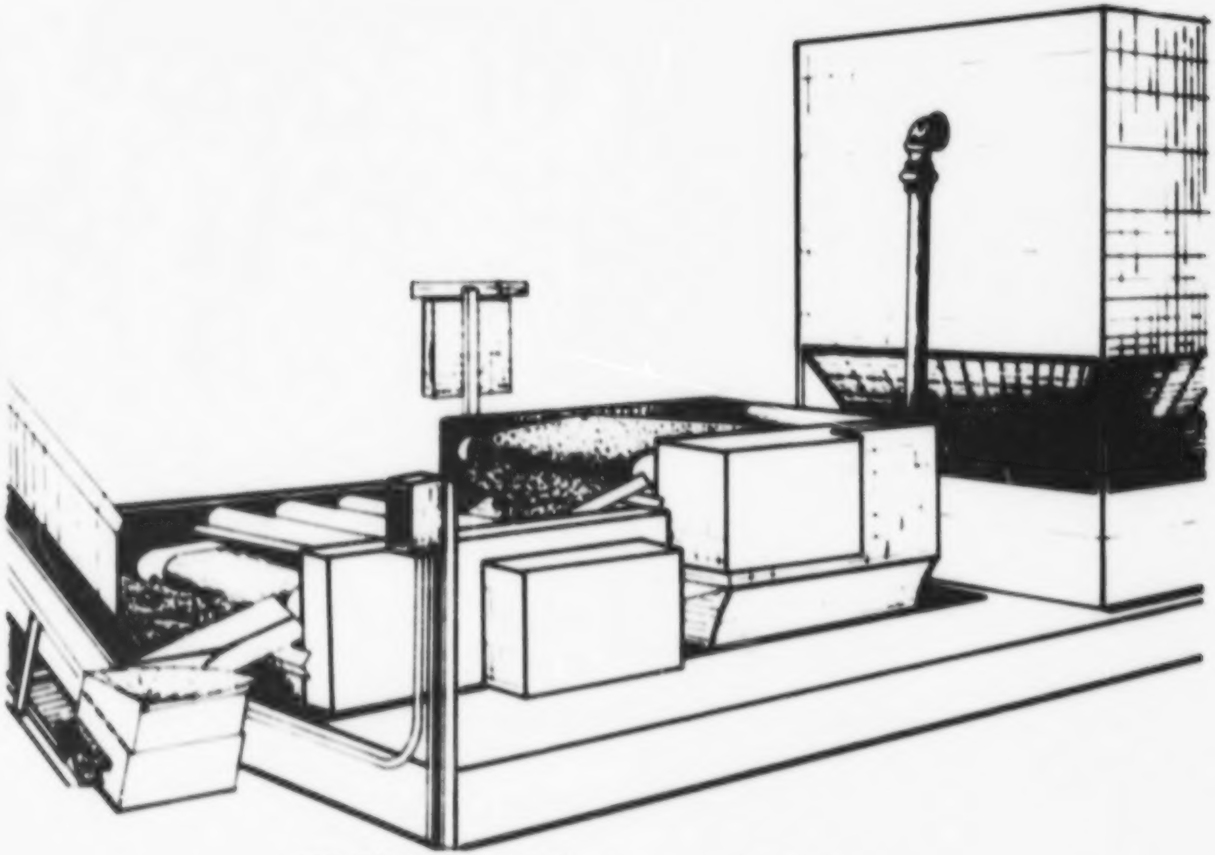


FIGURE 7-35. Moving screen concentrator system.

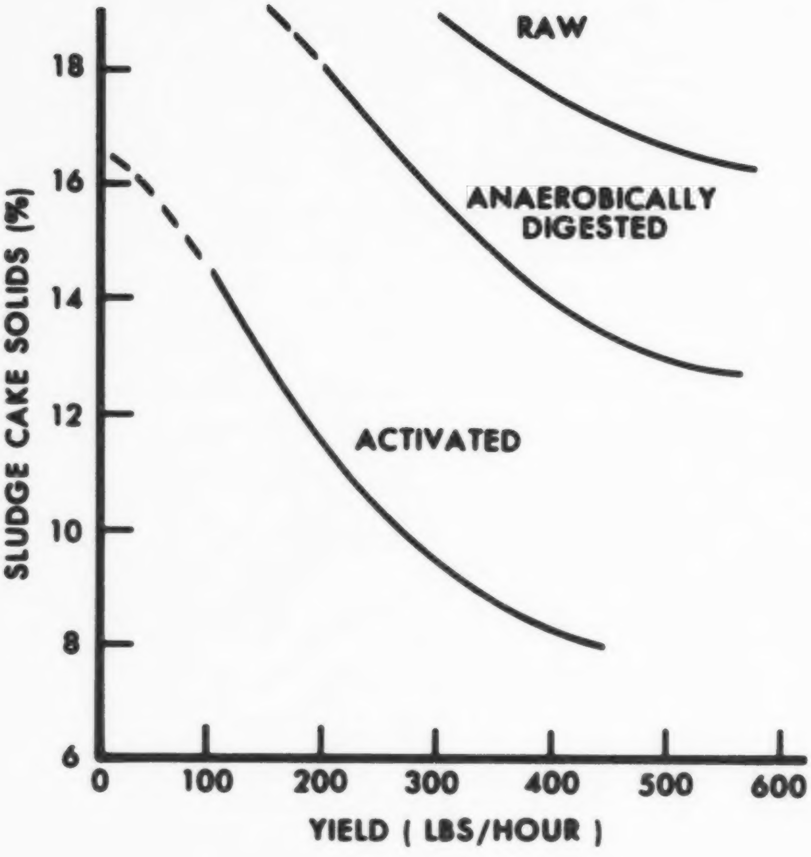


FIGURE 7-36 Moving belt concentrator yield vs. cake solids

doses were found to range from 5 to 15 pounds per ton of dry solids depending on the type and initial solids concentrations of the sludge being dewatered. Typical results for dewatering of activated sludge follow:

Feed Conditions			Results		
Type	% Solids	lb/hr	Conditioner $/ton*	% Cake Solids	% Solids Capture
EAS	0.7	298	6.60	8.8	90

*Based on 1970 costs data.

Total electrical costs for this unit were $0.16 per hour, and operator requirements were estimated at one hour per day [56]. This concentration of activated sludge could be achieved at an equivalent solids loading of about 19 psf/hour on a 40-inch wide belt [35]. An 80-inch model is also available, and it can process from 400 to 800 lb/hour of excess activated sludge and 800 to 1,600 lb/hour of primary sludge. Final cake solids of 20 to 30 percent have resulted with primary sludges. The capital cost of the 40-inch model is reported to be about $30,000 and it provides a capacity of 200 to 400 lb/hour of activated sludge and 400 to 800 lb/hour of primary sludge. The unit is normally supplied as a package which includes conditioning and thickening equipment. The chemical mixing and feed system is designed for easy operation and control. There are several installations of this system in operation or being planned. Its relatively low capital and power costs and adaptability to small plants are in its favor.

7.7.2 Belt Pressure Filters

Belt pressure filters are now being marketed by Passavant and Ralph B. Carter Co. The Carter Company introduced a new belt filter press to the U. S. in 1971. This system as shown schematically in Figure 7-37 was originally developed in Europe where it is widely used. An endless filter belt (a) runs over a drive and guide roller at each end (b,c) like a conveyor belt. The upper side of the filter belt is supported by several rollers (d). Above the filter bed a press belt (e) runs in the same direction and at the same speed, whose drive roller (f) is coupled with the drive roller (b) of the filter belt. The press belt can be pressed on the filter belt by means of a pressure roller system whose rollers (g) can be individually adjusted horizontally and vertically. The sludge which is to be dewatered is fed on the upper face of the filter belt and is continuously dewatered between the filter and press belts. After having passed the pressure zone, further dewatering in a reasonable time cannot be achieved by only applying static pressures. However, a superimposition of shear forces can effect this further dewatering. The supporting rollers of the filter belt and the pressure rollers of the pressure belt are adjusted in such a way that the belts and the sludge between them describe an S-shaped curve. Thus, there is a parallel

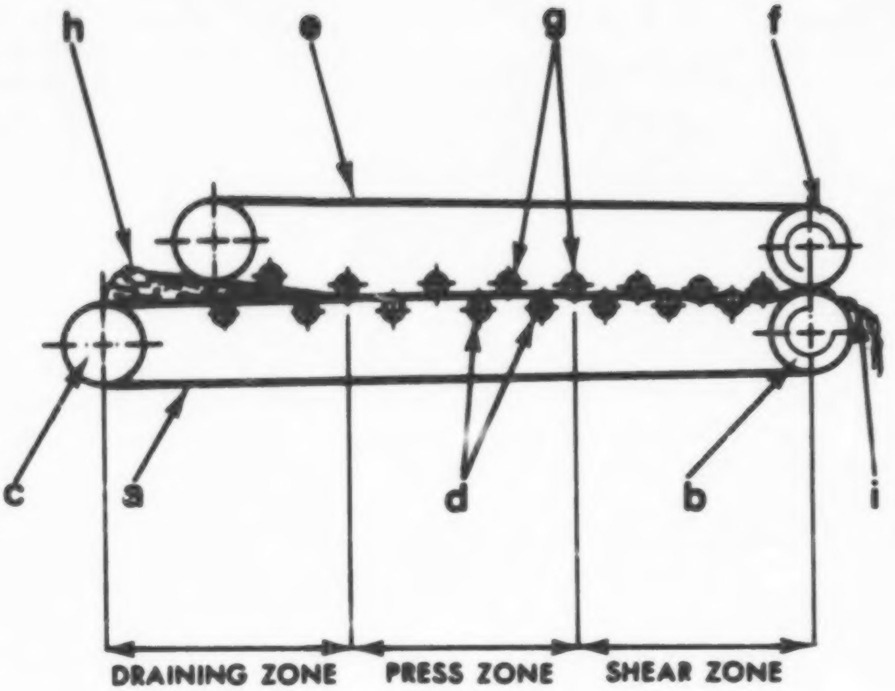

h
e
g
f
c
a
d
b
i
DRAINING ZONE
PRESS ZONE
SHEAR ZONE

FIGURE 7-37. Schematic construction of the belt filter press.

displacement of the belts relative to each other due to the difference in radius. After further dewatering in the shear zone, the sludge is removed by a scraper (i).

As can be seen, the belt filter press has three processing zones along the length of the unit. They are the initial draining zone, which is analogous to the action of a drying bed; the press zone, which involves application of pressure, and a shear zone in which shear is applied to the partially dewatered cake. The unit has recently been modified to a two-stage unit as depicted in the photograph in Figure 7-38 where the initial draining zone is on the top level followed by an additional section wherein pressing and shearing occur. A significant feature of the belt filter press is that it employs a coarse, mesh, relatively open weave, metal medium fabric. This is feasible because of the rapid and complete cake formation obtainable when proper flocculation is achieved.

The belt filter press, as well as the other systems described in this section, attempt to overcome the sludge pick up problem occasionally experienced with rotary vacuum filters. The belt filter press supplied by Carter is supplied in system form as shown in Figure 7-39. This system includes auxiliaries such as polymer solution preparation equipment, and automatic process controls. The Carter belt press has the capability of dewatering a digested combined primary and secondary sludge with an initial solids of 5.7 percent to a final cake solids concentration of 19 percent. It can do this at a rate of 6.7 lb/sq ft/hr and a chemical conditioning cost of $4.10/ton. Table 7-12 summarizes European installation as of 1971. This unit is currently being installed in several United States locations.

TABLE 7-12

EUROPEAN INSTALLATIONS OF THE BELT FILTER PRESSES

Type Sludge	Number of Plants	Population Equivalent
Primary/Biological	38	710,000
Primary/Biological/Chemical	4	132,000
Biological/Chemical	12	155,000
Industrial	13	

The Passavant belt filter press is somewhat similar to the Carter belt filter press, but includes the added feature of vacuum boxes in the free drainage zone. About 6 inches of

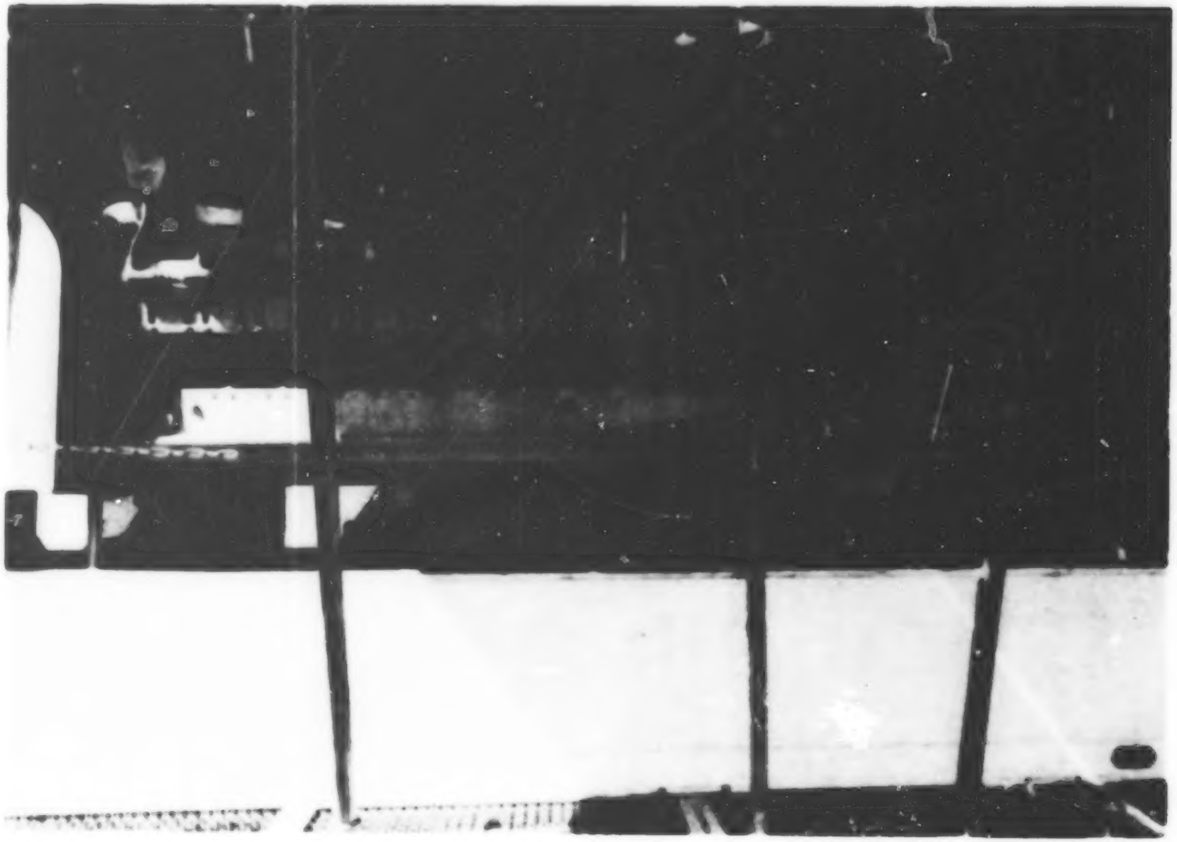


FIGURE 7-38 Belt filter press.

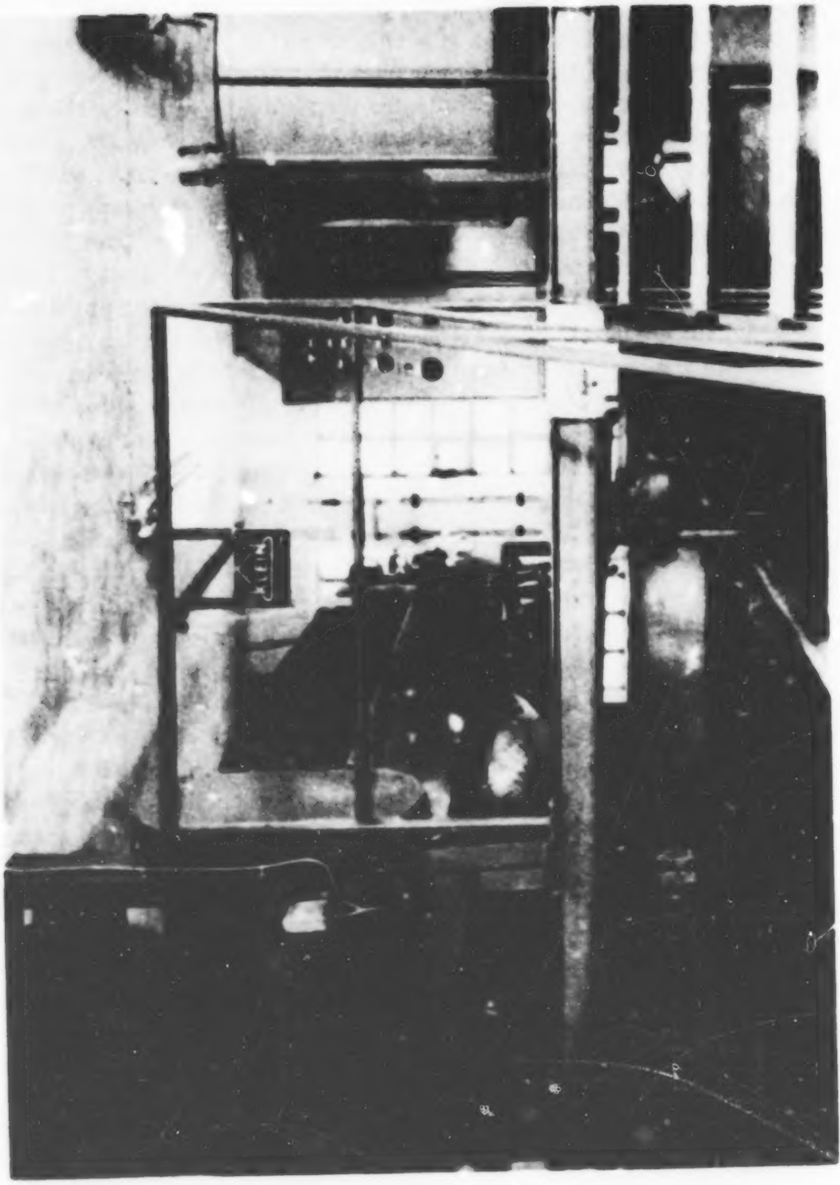


FIGURE 7-39. Belt filter press system.

vacuum (Hg) are applied to obtain higher cake solids. Figure 7-40 shows the initial test unit at Birmingham, Alabama. Passavant reports cake solids concentrations averaging 25 to 30 percent from the dewatering of mixed sludges.

7.7.3 Capillary Dewatering Systems

The Squeegee or capillary suction device is a unit offered by the Infilco Division of Degremont and is a new type of horizontal belt filter [57]. Figure 7-41 shows a drawing of a unit. The Squeegee incorporates a self-contained chemical feed system consisting of a variable rate diaphragm metering pump, mixer, and storage tank containing chemical coagulant. The sludge is metered by a variable rate sludge pump and flows to the unit prior to entrance into the influent box. The sludge is then distributed over the screen longitudinally through a series of openings which create a uniform level. This portion of the operation releases free water and increases the solids concentration level by approximately 25 percent. The free water released escapes through the screen and is collected in a trough for discharge. After the initial free water release, the screen carrier comes in contact with the capillary belt. The unusual feature of the Squeegee device is this capillary dewatering zone wherein the motive force for dewatering comes from the capillary action of the capillary belt. This belt is separated from the sludge by the carrier screen. The capillary dewatering zone is shown in Figure 7-42. Along the longitudinal path of the belt are a series of stations which extract the liquid from the saturated belt. One station is shown in Figure 7-43. The number and location of these stations can be varied on-site to optimize results depending on sludge characteristics. In these stations, the belt is momentarily separated from the screen and sludge, while the filtrate is removed from the belt. After the optimum amount of liquid has been extracted by the belt, the screen carrier and belt are again finally separated. The carrier and sludge continue along the longitudinal plane where a final compression zone extracts additional liquid for a final dehydration as shown in Figure 7-44. The sludge cake is then removed by a doctor blade in contact with the final compression roller and the cake drops onto a discharge chute. Before the carrier completes the cycle it passes through a high pressure washing station and returns for the next cycle. A belt washing station is also provided, which is controlled by a timer and solenoid valve that washes the belt at predetermined intervals.

Considerable data have been obtained on a pilot plant scale at the Long Road treatment plant near Pittsburg, Pennsylvania. A summary of this data appears in Table 7-13. Feed capacities from 2 to 4.5 lb/hr/ft^2 were achieved with the cake solids at discharge ranging from 15 to 18 percent for activated sludge. Machine capacities more than twice these values may be possible. It was found that the device can be operated without coagulant addition at a penalty to solids capture. With chemical addition of the polyelectrolyte at 10 lb/ton or about $4.00/ton a cake solids capture of 95 percent can be obtained when the machine is operated at 2.0 lb/hr/ft^2. Operating at higher machine capacities, i.e., 4.0 lb/hr/ft^2 and higher, it appears that the ferric chloride conditioner yields more economical system operation. Overall machine operation was found to depend on chemical addition,

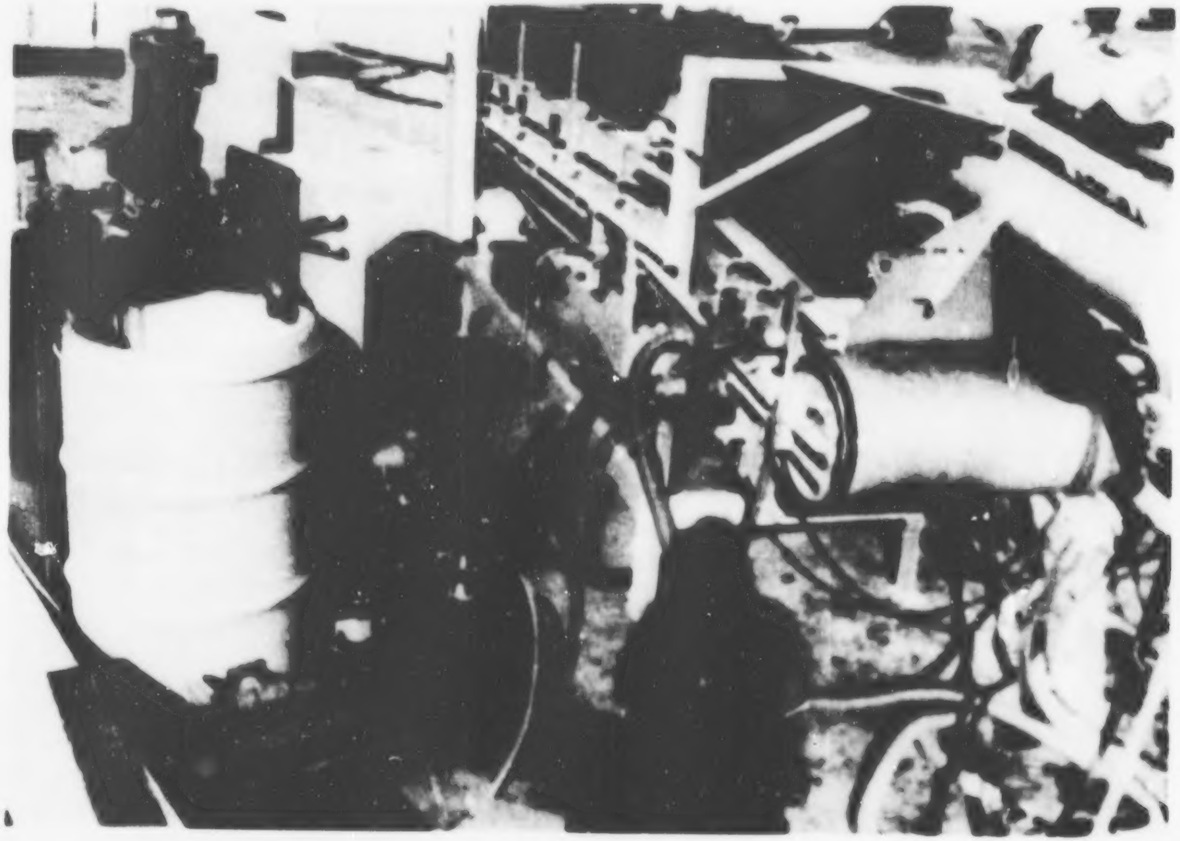


FIGURE 7-40 Passavant belt filter press.

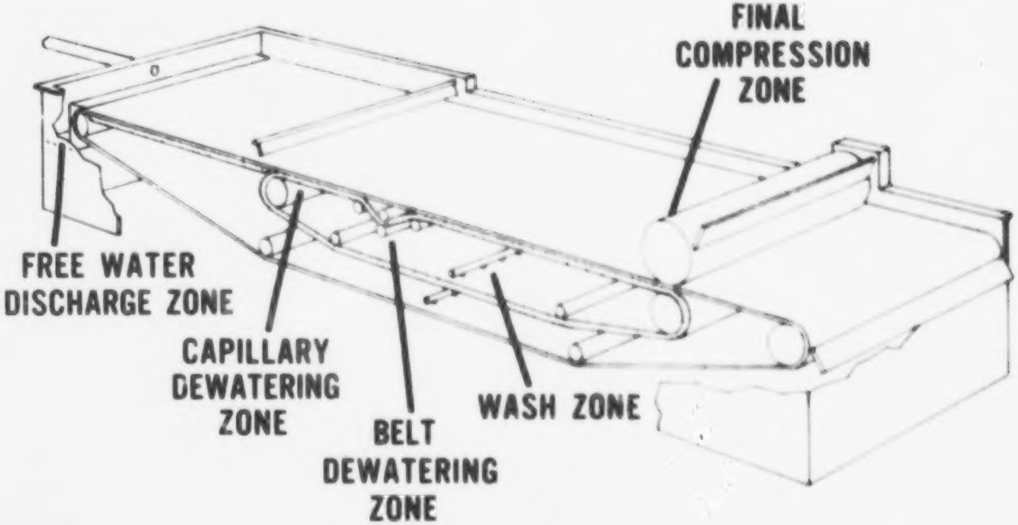

FINAL COMPRESSION ZONE
FREE WATER DISCHARGE ZONE
CAPILLARY DEWATERING ZONE
BELT DEWATERING ZONE
WASH ZONE

FIGURE 7-41. Squeegee capillary sludge dewatering unit

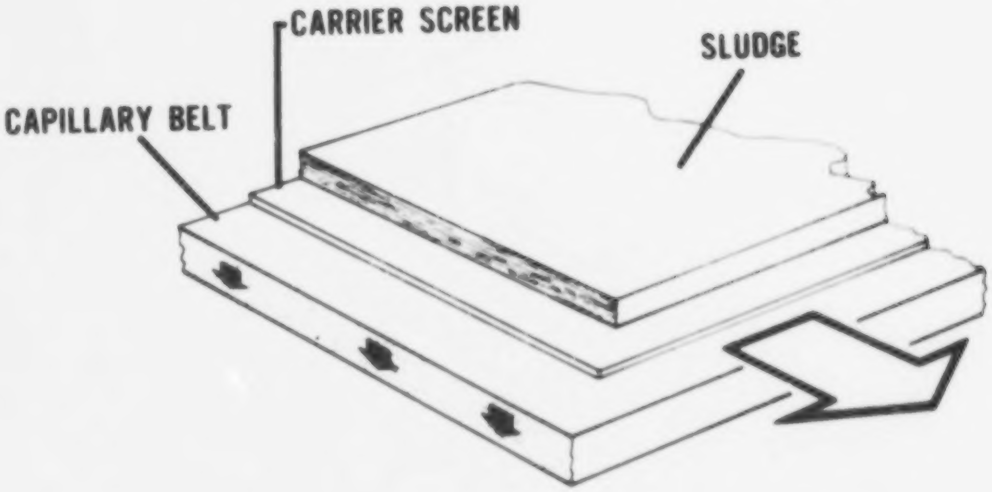

CARRIER SCREEN
SLUDGE
CAPILLARY BELT

FIGURE 7-42   Capillary dewatering zone

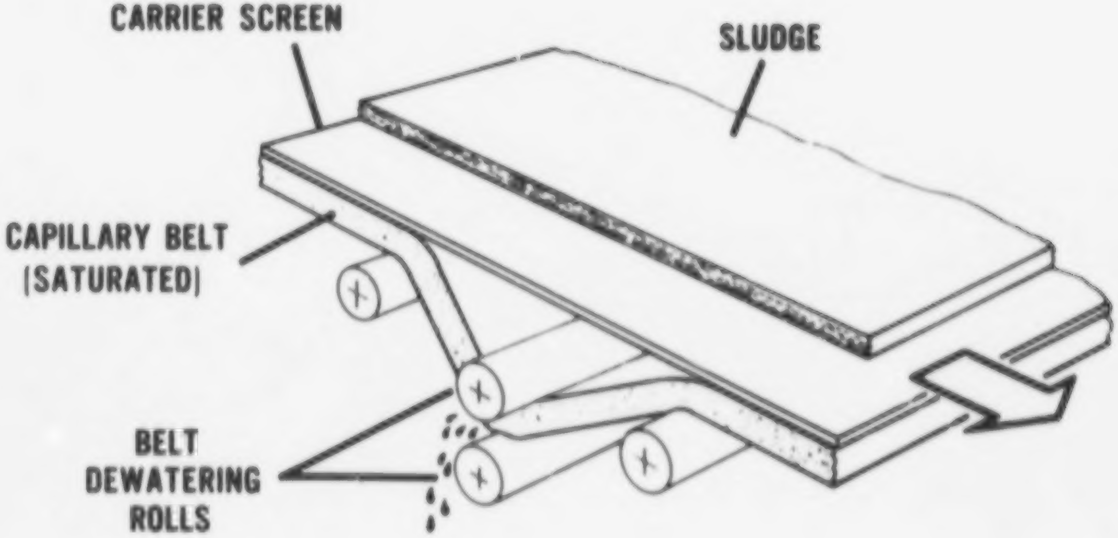

CARRIER SCREEN
SLUDGE
CAPILLARY BELT (SATURATED)
BELT DEWATERING ROLLS

FIGURE 7-43. Belt dewatering zone.

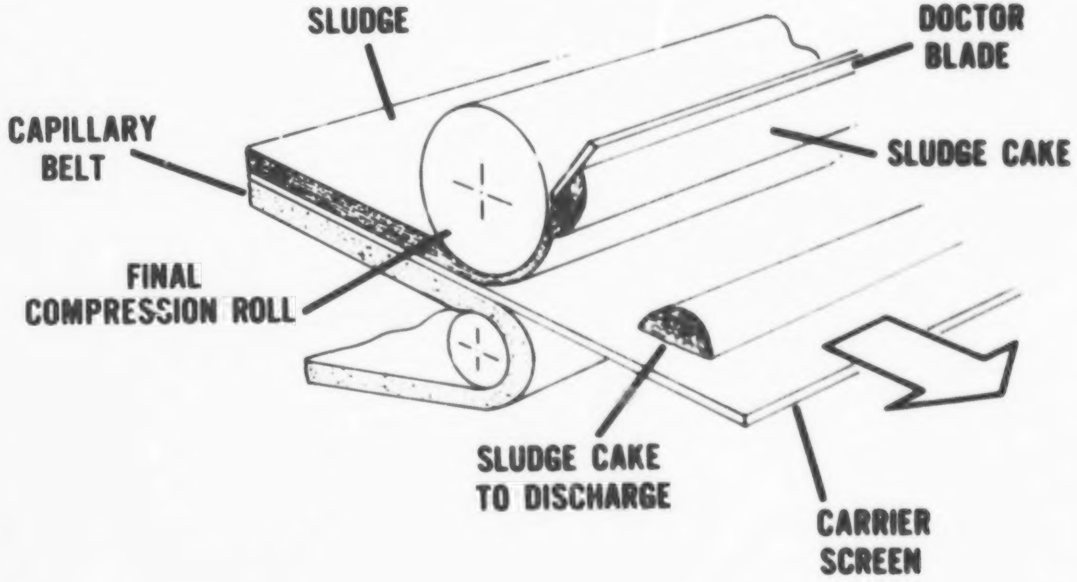

SLUDGE
DOCTOR BLADE
CAPILLARY BELT
SLUDGE CAKE
FINAL COMPRESSION ROLL
SLUDGE CAKE TO DISCHARGE
CARRIER SCREEN

FIGURE 7-44. Final compression zone.

sludge solids loading, and screen mesh size. Total cost estimates for capillary dewatering range from $19.36 to $39.67 per ton of dry sludge solids processed in plants of the two to four mgd size [58].

TABLE 7-13

SUMMARY OF PILOT PLANT CAPILLARY DEWATERING SYSTEM PERFORMANCE

Sludge Type	Machine Capacity (lb/hr/ft²)	Coagulant Cost ($/ton)*	Final Cake Solids (%)	Solids Recovery (%)
		Cationic Polyelectrolyte		
EAS	2.0	0	16-19	60-75
EAS	2.0	4.00	17-19	95
		Ferric Chloride		
EAS	3.0	0	16	50-65
EAS	4.3	6.40	15	91
		Cationic Polyelectrolyte		
EAS	4.5	4.00	14-15	80
		Ferric Chloride		
Digested	5.2	10.00	16-18	88
Mixed Primary	5.4	10.00	14-15	85

*Based on 1973 cost data.

Currently there is one plant scale unit in operation; and this is at Coral Springs, Florida, where an aerobically digested sludge is being processed. Reported results are similar to those obtained with the pilot unit an excess activated sludge. Another unit is scheduled to be installed at St. Charles, Illinois, in early 1975.

7.7.4 Rotating Gravity Concentration

This concept is employed in a unit marketed by the Permutit Co. under the trade name of "DCG sludge dewatering unit" and it is shown in Figure 7-45. At one time, Nichols Engineering offered a similar unit under the name "Roto-Plug" but the patent rights have since been sold to Permutit. The unit consists of two independent cells. The cells are formed by a fine mesh nylon filter cloth which travels continuously over front and rear guide wheels. The filter cloth is rotated by a drive roll and sprocket assembly, which also serves as the separator between the two cells. Dewatering occurs in the first cell, cake formation takes place in the second cell. Sludge is introduced in the dewatering cell, where initial liquid solids separation takes place. In this section liquid drains through the nylon filter cloth at a relatively high rate. The partially dewatered solids are then carried over the drive roll separator into the second cell or cake formation zone. Here they are continuously rolled and form into a cake of relatively low moisture content. The weight of this sludge cake presses additional water from the partially dewatered sludge. When the cake of dewatered solids grows to a certain size, excess quantities are discharged over the rim of the second cell to a conveyor belt which moves the material to a disposal point. Operation is continuous. Dewatering is accomplished entirely by gravity and without the application of either pressure or vacuum. Blinding tendencies are minimized, even on difficult-to-dewater sludges, because of the low differential pressure (1 inch or 2 inches) applied, and the tendency of sludge to adhere to itself and separate cleanly from the cloth.

The DCG unit is a concentrating device and, when more complete dewatering is required, a multiroll press is provided as illustrated in Figure 7-46. The MRP dewatering press consists of dual endless belts. Sludge cake, concentrated by the DCG, is fed by rotating blades to the space between the belts and graduated pressure is applied by the rollers to squeeze additional moisture through the cloth into the grooved support belt and thence to the drip pan. The dewatered cake is carried by the bottom cloth to the discharge point.

The general capabilities of the DCG and the MRP units are shown in Table 7-14 [59]. The dewatering belts in the MRP unit are reported to have a life of about 6 months. The first United States installation was made in 1962 at Caldwell, New Jersey and by 1969, 17 installations had been made [60]. Most of the installations were on aerobically digested sludges. Production of a 25 percent cake from a 6 percent feed of raw primary sludge without conditioning has been reported in one case [61]. The initial plant at Caldwell, New Jersey reports polymer costs of $8 to $10 per ton when operating on an anaerobically digested primary plus trickling filter humus sludge of 4 to 5 percent solids concentration and producing 15 percent cake solids [62,59].

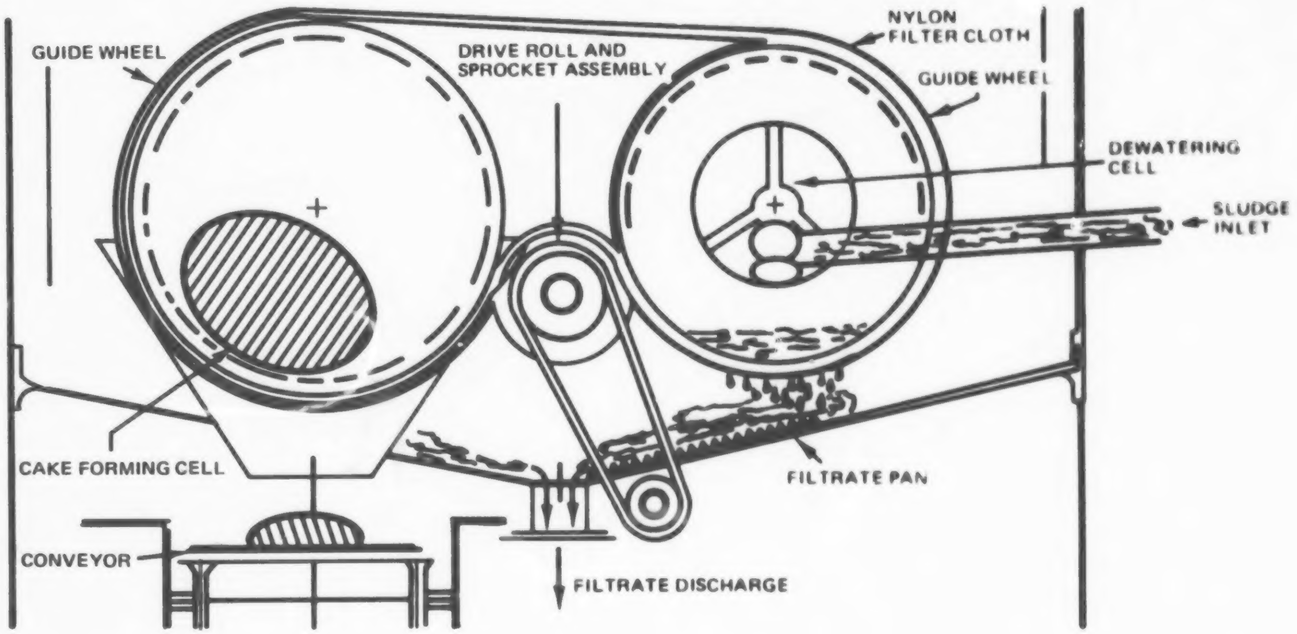

GUIDE WHEEL
DRIVE ROLL AND SPROCKET ASSEMBLY
NYLON FILTER CLOTH
GUIDE WHEEL
DEWATERING CELL
SLUDGE INLET
CAKE FORMING CELL
FILTRATE PAN
CONVEYOR
FILTRATE DISCHARGE

FIGURE 7-45. Rotating gravity concentrator.

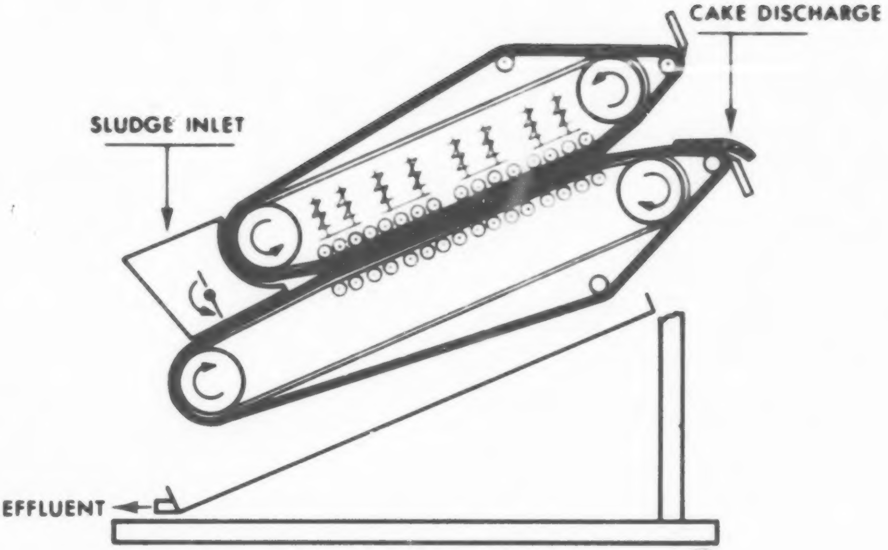

CAKE DISCHARGE
SLUDGE INLET
EFFLUENT

FIGURE 7-46. Schematic of MRP section.

TABLE 7-14

CAPABILITIES OF THE MRP AND DCG UNITS

Type Sludge	% Dry Solids Influent	 DCG	 MRP
Raw Primary	3	12-17	20-23
Anaerobic Digested	4	12-15	18-20
Aerobic Digested	1.5-3.0	10-12	16-20
Excess Activated	1.9-3.0	9.4-13	18-23
Oxygen Activated	3	10	16

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CHAPTER 8

SLUDGE REDUCTION

8.1 Methods, Functions, and Occurrences

Sludge reduction processes are generally thermal ones and provide a major reduction in the sludge solids. Although heat drying is not truly a reduction process, it occupies the same relative position in the sequence of sludge processing as major reduction processes. It is included in this chapter for purposes of simplification. Established and experimental sludge reduction processes are listed and categorized in Table 8-1.

TABLE 8-1

REDUCTION PROCESSES

Reduction Process	Pretreatment Required	Additional Processing Requirements
Established Processes		
Incineration	Thickening and Dewatering	Landfill ash
Wet Air Oxidation	Thickening	Treat cooking liquor, Landfill ash
Heat Drying	Thickening and Dewatering	Use dried sludge as soil conditioner
Experimental Processes		
Pyrolysis	Thickening	Utilize by-products of gas, carbon, steam. Dispose of residue
Incineration/Power or steam generation	Thickening and Dewatering	Landfill ash

Where sufficient land was available for proper disposal of liquid, or more frequently, dewatered sludge cake, past practice at wastewater treatment plants has been to omit incineration or other reduction processes due to the cost. With decreasing land availability and the possibility of more stringent standards for land disposal, reduction processes have been receiving revived attention. This is also true of municipal solid waste disposal practices. The concurrent increase in the price of energy has sparked wide interest in the combined incineration of solid waste residues and wastewater sludges in reduction systems incorporating energy recovery as a prime feature.

8.2 Incineration

The basic elements of sludge incineration are shown schematically in Figure 8-1. An incinerator is usually part of a sludge treatment system which includes sludge thickening, a macerating or disentegrating system, a dewatering device (such as a vacuum filter, centrifuge, or filter press), an incinerator feed system, air pollution control devices, ash handling facilities, and the related automatic controls. Important considerations in evaluating incineration methods include the composition of the sludge feed and the amount of auxiliary fuel required. Air pollution constraints and resultant equipment and treatment requirements as well as ash disposal are also important.

8.2.1 Composition of Sludge Feed

A primary consideration in the cost-effectiveness of sludge incineration is the effect of sludge feed composition on auxiliary fuel requirements. Other variables of importance are the type of incinerator employed, excess air requirements, operating temperatures necessary for odor control and other air pollution constraints. Heat yield from a given sludge is a function of the relative amounts and elemental composition of the contained combustible elements.

Wastewater sludges have been incorrectly classified as low grade fuels. While the heating value of the dried sludge solids may be substantial, sludges cannot be considered as fuel because of the high moisture content of the raw sludge. Sludges contain significant amounts of the three principal combustible elements: carbon, hydrogen, and sulfur. It is possible, using the Dulong formula [1], to estimate calorific value of sludges from knowledge of the amounts of these three elements in a given sludge. However, experience has shown that results so calculated are very often inaccurate, so that the only reliable method of determining heat value is to carry out calorimeter tests. Inert and moisture contents of the feed sludge also affect the heat requirements for complete combustion. In the past, the feed sludge's dry solids content has been overemphasized, while insufficient regard has been shown for the effect of the sludge's chemical composition and inert content on auxiliary fuel requirements. The anaerobic digestion and heat treatment processes reduce the volatile content and increase the inert noncombustible content with the resultant lower fuel value for a sludge. Physical chemical sludges also have low heat

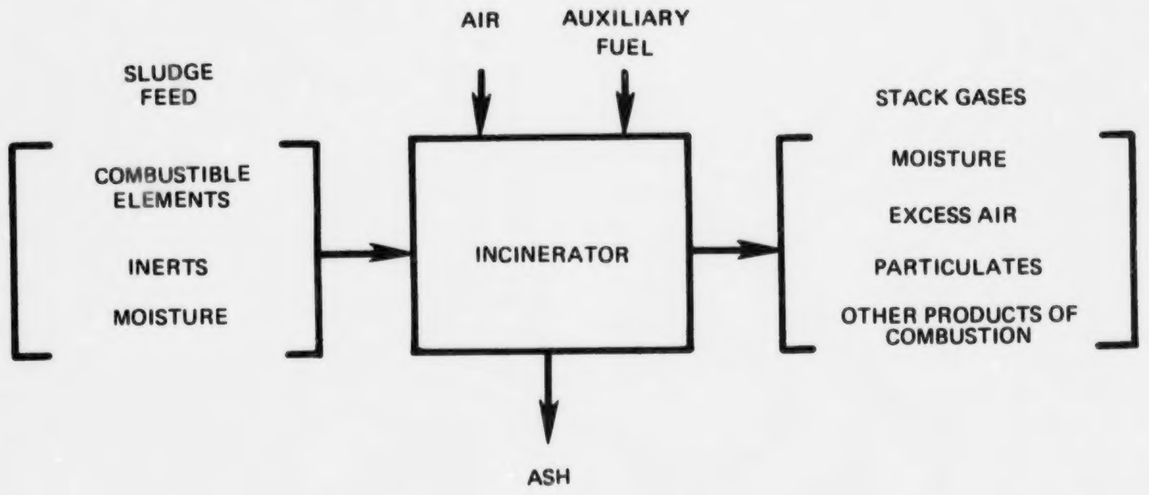

AIR
AUXILIARY FUEL
SLUDGE FEED
COMBUSTIBLE ELEMENTS
INERTS
MOISTURE
INCINERATOR
ASH
STACK GASES
MOISTURE
EXCESS AIR
PARTICULATES
OTHER PRODUCTS OF COMBUSTION

FIGURE 8-1. Sludge incineration.

values due to inert content. Table 8-2 gives the heat value of various sludge types. Table 8-3 gives some representative heating values of various sludge constituents. Pretreatment methods such as chemical conditioning and dewatering do result in substantial reduction of incineration fuel requirements, but frequently they do so by creating increased energy demands on other unit processes.

TABLE 8-2

EFFECTS OF PRIOR PROCESSES ON FUEL VALUE [1]

Type Sludge	Calorific Value (BTU/lb of dry solids)
Raw Primary	9,500
Anaerobically Digested Primary	5,500
Raw (Chem. Precip.) Primary	7,010

TABLE 8-3

REPRESENTATIVE HEATING VALUES OF SOME SLUDGE MATERIALS [1]

Material	Combustibles (%)	Heating Value (BTU/lb of dry solids)
Grease and Scum	88	16,700
Raw Wastewater Solids	74	10,300
Fine Screenings	86	9,000
Ground Garbage	85	8,200
Digested Sludge	60	5,300
Chemical Precipitated Solids	57	7,500
Grit	33	4,000

8.2.2 The Incineration Process

Incineration is a two-step process involving drying and combustion. In addition to fuel and air; time, temperature, and turbulence are necessary for a complete reaction. The drying step should not be confused with preliminary dewatering; this dewatering is usually by mechanical means and precedes the incineration process in most systems. A sludge with a moisture content of about 75 percent is delivered to the incinerators. Since a typical sludge might contain 3 pounds of water for each pound of dry solids and have a volatility of 75 percent, the heat required to evaporate the water nearly balances the available heat from combustion of the dry solids.

Drying and combustion may be done in separate units or successively in the same unit. Manufacturers have developed diversified types of equipment. The two major incineration systems employed in the United States are the multiple hearth furnace and the fluidized bed incinerator. The drying and combustion process consists of the following phases: (a) raising the temperature of the feed sludge to 212° F, (b) evaporating water from the sludge, (c) increasing the water vapor and air temperature of the gas, and (d) increasing the temperature of the dried sludge volatiles to the ignition point. Practical operation of an incinerator requires that air in excess of theoretical requirements be supplied for complete combustion of the fuel. The introduction of excess air has the effect of reducing the burning temperature and increasing the heat losses from the furnace.

Heat is emitted by the burning of sludge in a furnace. Some of this heat is absorbed by the furnace and lost by radiation. A large portion of the emitted heat is lost with the stack gases, while a small portion is lost with the ash. The difference between the heat generated and the heat lost is available for heating the incoming sludge and air. Self-sustained combustion is often possible with dewatered raw sludges once the burning of auxiliary fuel raises incinerator temperatures to the ignition point.

The combustion process requires, in addition to fuel and air the proper amount of time, temperature, and turbulence for a complete reaction [2]. The reciprocal relationship between time and temperature which is noted in many chemical processes is also of importance in combustion. A surprisingly large part of the heat produced in the combustion reaction (about 50 percent) must be delivered to the entering fuel and air to keep the reaction going. This is less surprising when it is realized that the fuel and air must be heated to a relatively high ignition temperature to maintain the process. An important feature of burner and furnace design is the optimum combining of the required ignition energy (heat) and the fuel and air. In the design of burners for gaseous or liquid fuels, turbulence plays an important part in the achievement of ignition temperature. In the design of equipment currently used for incineration, turbulence is a less important factor and heating of the sludge to ignition temperature is usually achieved by radiation from the hot refractory surfaces. There are a number of variables which influence the amount of fuel required and the resulting cost for sludge incineration. Principal variables are the moisture and volatile solids content of the sludge, and their effect on the sludge's heat content is shown in Figure 8-2. The importance of obtaining the maximum solids concentration is illustrated by Figure 8-3. For example, at 25 percent total solids there is only enough heat available in this particular sludge to raise the combustion products and moisture to 900° F and this temperature is far below the accepted 1350 to 1400° F necessary for deodorizing the stack gases of a conventional combustion unit.

Excess air added to the combustion chamber increases the opportunity of contact between fuel and oxygen and this is essential for combustion to proceed. To insure complete thermal oxidation, it has been necessary to maintain 50 to 100 percent excess air over the stoichiometric amount of air required in the combustion zone. This excess air is undesirable because it quenches the reaction temperature by pirating 12 to 24 percent of the input BTU's for heating of the excess air. If excess air is not supplied for this reason, it is difficult to maintain the minimum deodorizing temperature. Therefore, a closely controlled minimum excess air flow is desirable for maximum thermal economy. The amount of excess air required varies with the type of incineration equipment, the nature of the sludge to be incinerated, and the disposition of the stack gases. The impact of use of excess air on the cost of fuel in sludge incineration is shown in Figure 8-4. When the amount of excess air is inadequate, only partial combustion of carbon occurs, and carbon monoxide, soot, and odorous hydrocarbons are formed. Further, the heat recovered from the partial burning of the carbon is substantially reduced, since the heat value of carbon monoxide is only 4,400 BTU/lb.

Preheating of air is an important step in improving the thermal economy. Air preheat affords an increase in capacity for a given size reactor since the combustion gas volume is used more effectively and auxiliary fuel requirements are reduced. However, it should be noted that the preheat exchanger has a significant capital cost and should only be recommended after a complete economic evaluation of the process.

8.2.3 Analysis of Incineration Processes

An effective incineration system analysis should include material and heat balances. Cost estimates should consider the potential incineration system's effects on other unit processes. A typical material balance around a fluid bed incinerator is shown in Figure 8-5. Since reasonably accurate determination of the sludge input analysis is essential, the need for sludge process evaluation work and sampling is apparent. Long-term wastewater treatment objectives are essential for the proper sizing of an incinerator [5].

Owen [1] has presented a detailed example of a heat balance for a multiple hearth furnace incinerating a fresh primary sludge with a volatile content of 70 percent from the Ashland, Ohio, wastewater treatment plant. The heat balance comprised the following quantities:

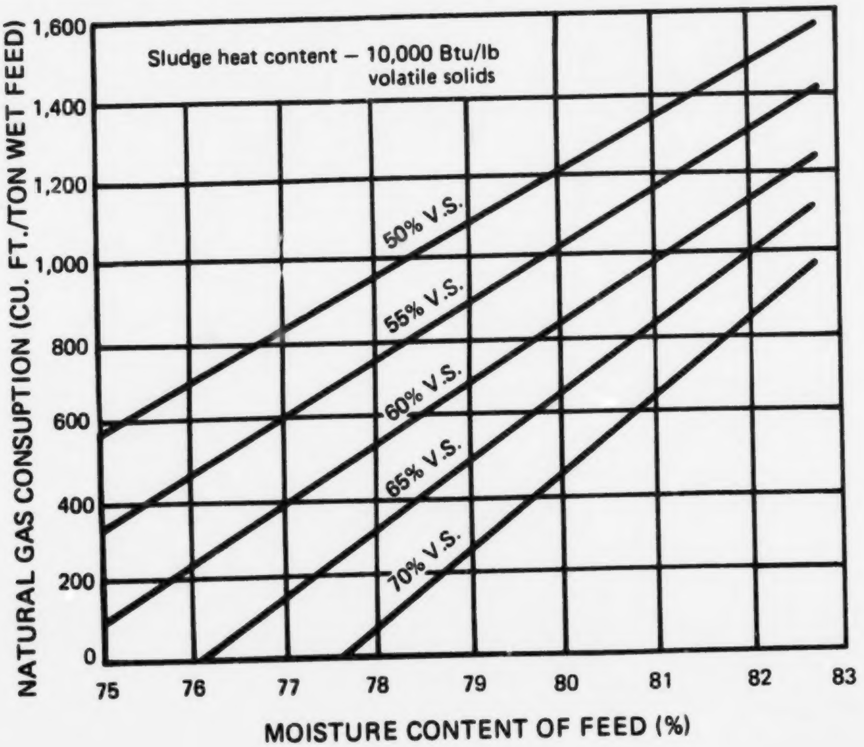


FIGURE 8-2. The effects of sludge moisture and volatile solids content on gas consumption.

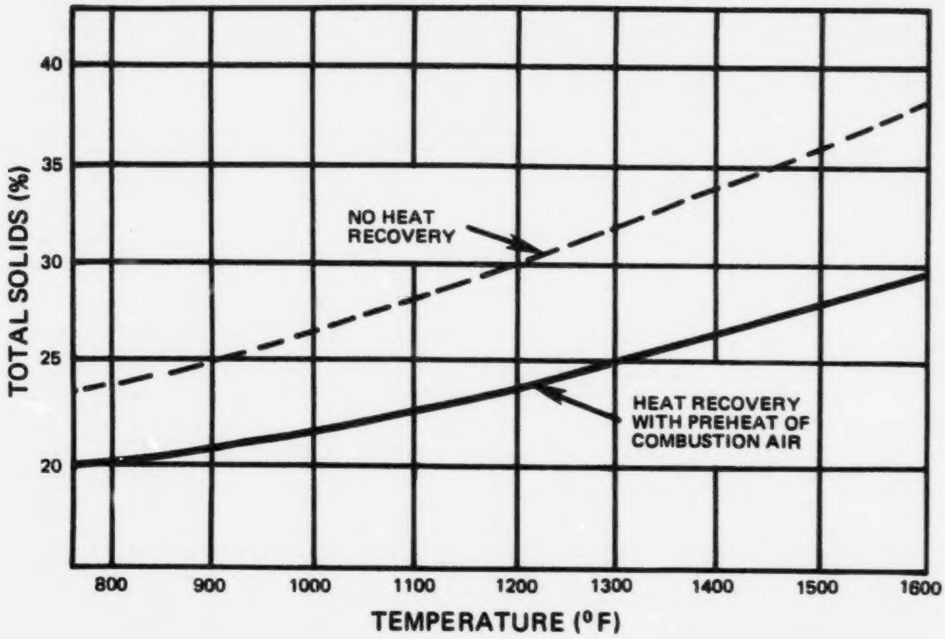


FIGURE 8-3. Equalibrium curves relating combustion temperatures to cake concentration [3].

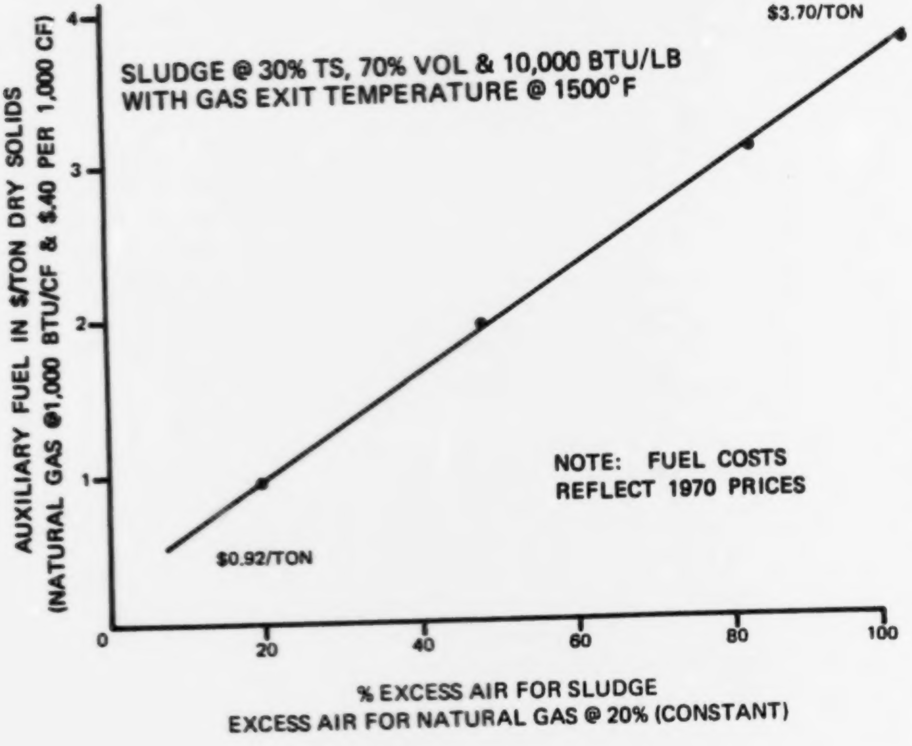


FIGURE 8-4. Impact of excess air on the cost of natural gas in sludge incineration [3].

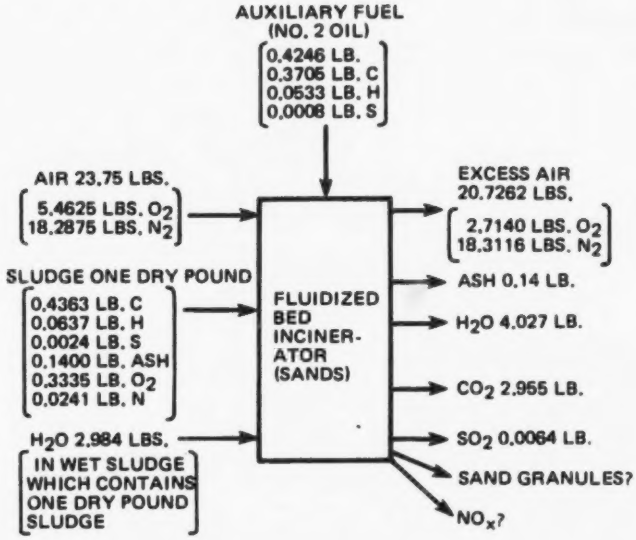

AUXILIARY FUEL
(NO. 2 OIL)
0.4246 LB.
0.3705 LB. C
0.0533 LB. H
0.0008 LB. S
AIR 23.75 LBS.
5.4625 LBS. O_2
18.2875 LBS. N_2
SLUDGE ONE DRY POUND
0.4363 LB. C
0.0637 LB. H
0.0024 LB. S
0.1400 LB. ASH
0.3335 LB. O_2
0.0241 LB. N
H_2O 2.984 LBS.
IN WET SLUDGE
WHICH CONTAINS
ONE DRY POUND
SLUDGE
FLUIDIZED
BED
INCINER-
ATOR
(SANDS)
EXCESS AIR
20.7262 LBS.
2.7140 LBS. O_2
18.3116 LBS. N_2
ASH 0.14 LB.
H_2O 4.027 LB.
CO_2 2.955 LB.
SO_2 0.0064 LB.
SAND GRANULES?
NO_x?

FIGURE 8-5. Material balance for fluidized bed sewage sludge incineration [4].

- Heat absorbed by
  - Latent heat in free moisture and moisture of combustion
  - Sensible heat in gases of combustion and excess air
  - Ash
  - Radiation
  - Shaft cooling air
- Heat evolved from
  - Combustibles in
    - Solids
    - Fuel

In a heat balance, the total heat absorbed must equal the total heat evolved.

8.2.4 Multiple Hearth Incineration

The multiple hearth furnace is the most widely used wastewater sludge incinerator in the United States today, because it is simple, durable, and has the flexibility of burning a wide variety of materials even with fluctuations in the feed rate. There were about 120 of these units installed for wastewater sludge combustion, as of 1970 [3]. A typical multiple hearth furnace is shown in Figure 8-6 and consists of a circular steel shell surrounding a number of solid refractory hearths and a central rotating shaft to which rabble arms are attached. Since the operating capacity of these furnaces is related to the total area of the enclosed hearths, they are designed with diameters ranging from 54 inches to 21 ft 6 inches and from four to eleven hearths. Capacities of multiple hearth furnaces vary from 200 to 8,000 lb/hr of dry sludge with operating temperatures of 1,700° F. The dewatered sludge enters at the top through a flapgate and proceeds downward through the furnace from hearth to hearth through the rotary action of the rabble arms.

The hearths are constructed of high heat duty fire brick and special fire brick shapes. The upper or No. 1 hearth in furnaces having an even number of hearths has a central opening or port through which sludge passes to the second hearth. The upper hearth in this case is termed an in-feed hearth; and the second, or next lower, an out-feed hearth. The top, or No. 1 hearth of furnaces having an odd number of hearths, is an out-feed hearth; and the next lower, or No. 2, is an in-feed hearth. Out-feed hearths have ports or drop-holes around the periphery of the hearth through which the sludge passes to the next lower

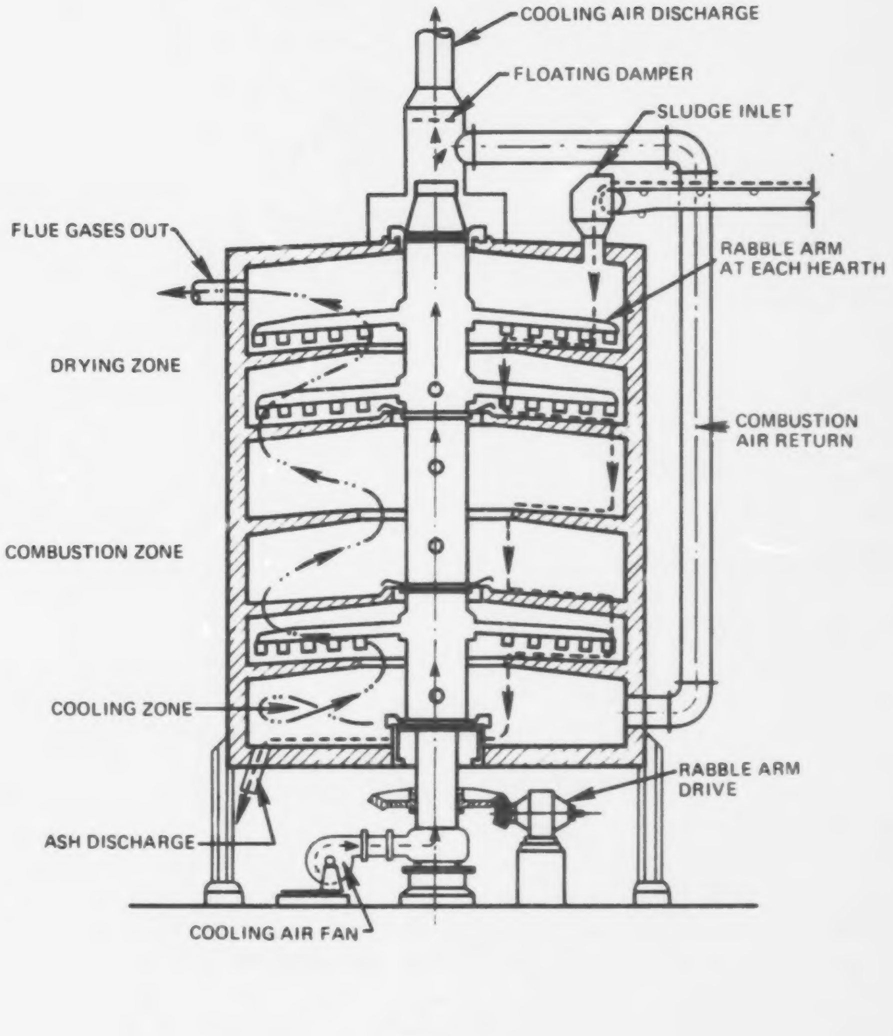

COOLING AIR DISCHARGE
FLOATING DAMPER
SLUDGE INLET
FLUE GASES OUT
RABBLE ARM AT EACH HEARTH
DRYING ZONE
COMBUSTION AIR RETURN
COMBUSTION ZONE
COOLING ZONE
RABBLE ARM DRIVE
ASH DISCHARGE
COOLING AIR FAN

FIGURE 8-6. Cross section of a typical multiple hearth incinerator.

in-feed hearth. The central circular opening of the in-feed hearths are constructed to leave a large clear opening between the edge of the hearth and the wall of the shaft. The ports of the out-feed hearth are constructed to provide openings all around the hearth and permit a well distributed supply of sludge to drop to the next lower hearth. These openings tend to regulate gas velocities also.

Two doors are generally provided at each hearth. They are fitted to cast iron frames and have machined faces to provide reasonably tight closures. An observation port with closure is provided in each door. Since the furnace may operate at temperatures up to 2,000° F, the central shaft and rabble arms are effectively cooled by air supplied in regulated quantity and pressure from a blower which discharges air into a housing at the bottom of the shaft. The central shaft is an iron column cast in sections. The sections enclosed by the furnace have a tubular inner column called the cold air tube. The annular space between the inner tube and the outer wall of the shaft exposed to furnace heat, serves as a passageway for hot air and is referred to as the hot air compartment. The shaft is motor driven through means permitting adjustment of speeds of from about one-half to one and one-half revolutions per minute. Two or more rabble arms are connected to machined arm sockets in the shaft at each hearth. Each rabble arm is constructed with a central tube for the purpose of conducting air from the cold air tube of the central shaft to the extreme end of the rabble arm, thence back through an outer air space in the arm to the shaft and through passages opening into the hot air compartment of the central shaft. The air may be discharged to atmosphere or returned to the bottom hearth of the furnace as preheated air, for combustion purposes.

The rabble arms provide mixing action as well as rotary and downward movement of the sludge. The flow of combustion air is countercurrent to that of the sludge. Gas or oil burners are provided on some of the hearths for furnishing heat for start-up or supplemental use as required. As shown in Figure 8-7, three phases of the incineration process occur at different levels in the furnace. In its travel across the hearths, sludge is constantly turned and broken into smaller particles by the rotating rabble arms. Thus a maximum sludge surface is exposed to the hot furnace gases which induces rapid and complete drying as well as burning of sludge. The rabble arms also form spiral ridges of sludge on each hearth, and the surface area of these ridges varies with the angle of repose of the sludge. This angle varies with the moisture of the material. In any case, the surface area of sludge exposed to the hot gases is considerably greater than the hearth area. Thus, an effective area of as much as 130 percent of the hearth area is produced. While the rabble arms provide significant solids-gas contact time on the hearths, the overall contact time is actually still greater, due to the fall of the sludge from in-hearth and out-hearth ports through the countercurrent flow of hot gases.

When burning sludge or any similar fuel, a measure of the work done by the heat energy in evaporating moisture is given by the drop in temperature of the hot gases as they pass between the combustion zone and the gas outlet of the furnace. The temperature of the combustion gases in a boiler furnace may average 2,000° F, but by giving up heat to the

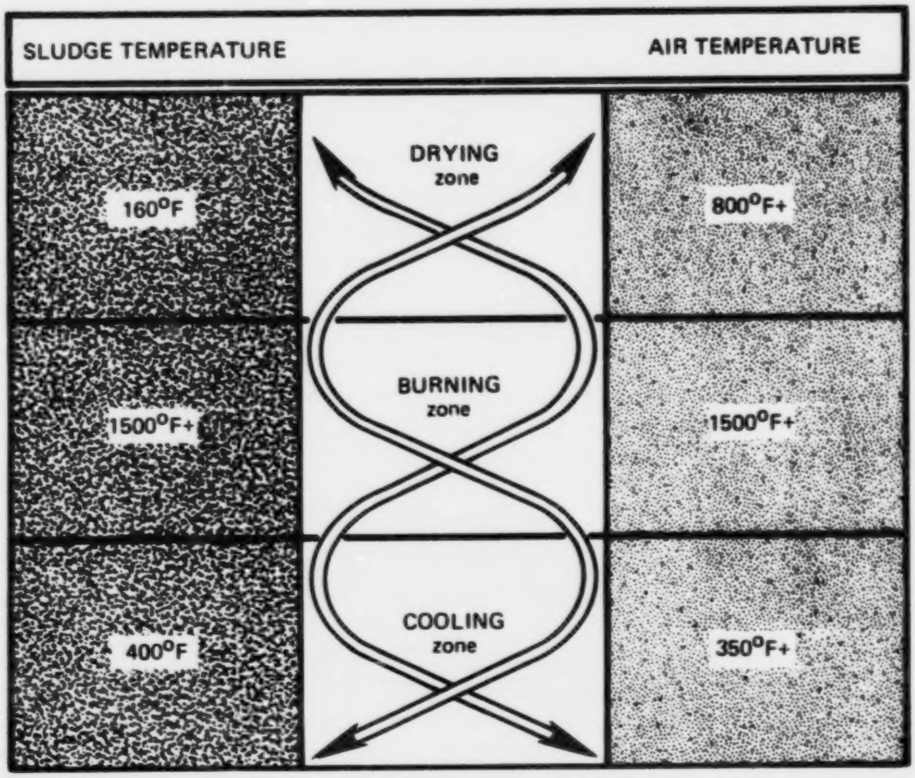

SLUDGE TEMPERATURE
AIR TEMPERATURE
160°F
DRYING zone
800°F+
1500°F+
BURNING zone
1500°F+
400°F
COOLING zone
350°F+

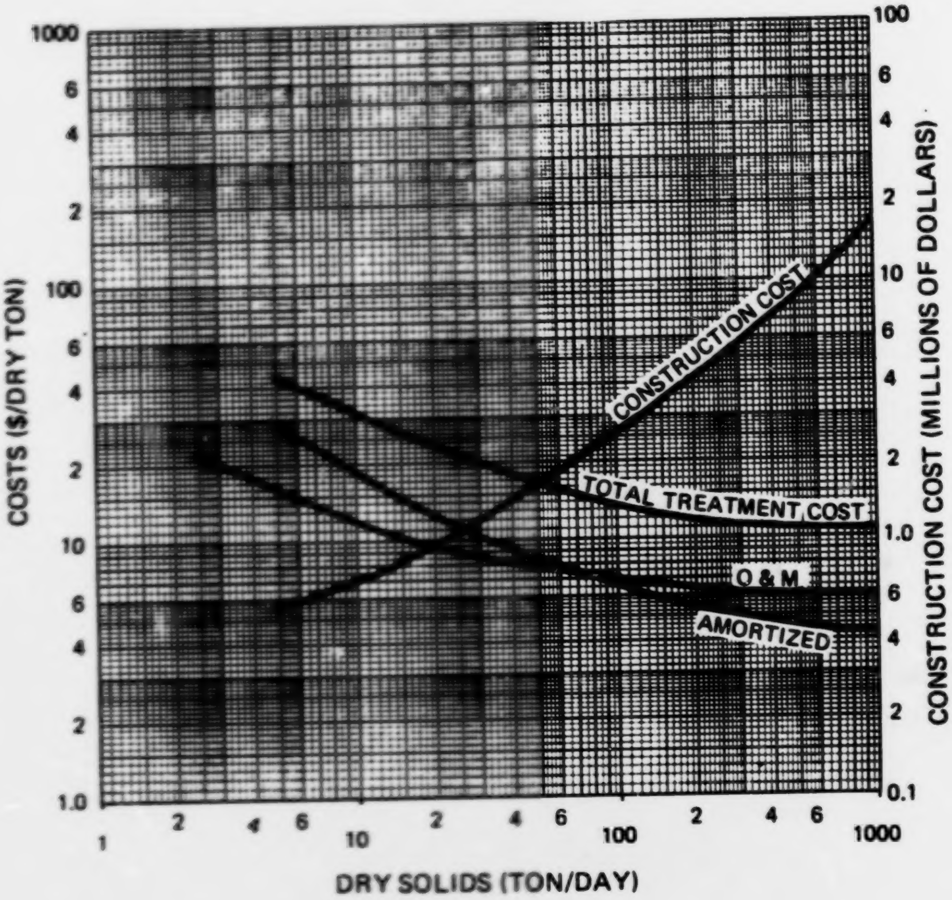
FIGURE 8-7. Multiple hearth process zones.

boiler, may leave the furnace at 500° F. Similarly, a multiple hearth sludge furnace may generate gas temperatures exceeding 1,500° F in the combustion zone. These gases sweep over the wet, cold sludge in the drying zone and perform useful work by giving up a considerable portion of their heat for evaporation of moisture. In this heat exchange, the gas temperature may drop as low as 500° F at the gas outlet. But while this exchange of heat evaporates an important percentage of sludge moisture, it does not raise the sludge temperature higher than about 160° F because the evaporation of water cools the mass it leaves. Since no significant quantity of odoriferous matter is distilled, exhaust gases need not be raised in temperature in an afterburner to destroy odors. Distillation of odoriferous material from sludge containing 75 percent moisture does not occur until 80 to 90 percent of the water has been driven off and, by this time, the sludge is down far enough in the incinerator to encounter gases hot enough to burn the odoriferous materials. Generally speaking, when fuel is required to maintain combustion in a multiple hearth furnace, a gas outlet temperature above 900° F indicates more fuel than required for incineration is being burned.

Some incinerator installations do provide high temperature treatment of the stack gases. The method is used is to convey the gases to a chamber where the temperature is raised by burning auxiliary fuel in direct contact with the gases before venting to the atmosphere. The cost of fuel for such a unit is substantial and can increase the *total cost* of sludge incineration by 10 percent.

The multiple hearth system can be provided with an instrumentation system which will convey the critical operating data to a central control panel. Temperature data can be recorded for each hearth, cooling, and exhaust, and scrubber inlet gas. The temperature can be controlled on each hearth to within ± 40° F. Malfunctions such as burner shutdown, furnace overtemperature, draft loss and feed belt shutdown can be monitored. In the event of power or fuel failure, the furnace should be shut down automatically and the shaft cooling air fan automatically transferred to a standby power source to avoid melting the shaft which drives the rabble arms. Problems encountered with multiple hearth furnaces have included: (a) failure of rabble arms and teeth, (b) failure of hearths, and (c) failure of refractories. Improvements in materials used in constructing the rabble arms and teeth have reduced the first problem by providing a greater margin between the yield point of the materials used and the normal operating temperatures in the furnace. Many refractory problems result from the need for careful heating and cooling of the furnaces during start-up and shutdown. From 24 to 30 hours are required to bring the furnace up to temperature or to cool it. This is an operational disadvantage and a procedure, which if not carefully followed, can contribute to refractory problems. However, there are several installations which do operate intermittently without significant refractory problems. Some difficulty has also been encountered with the ultraviolet scanners which are used to monitor loss of flame on a hearth since needless shutdowns of the furnace can result from a scanner malfunction.

As with all unit process costs, the cost of multiple hearth incineration varies with plant size and this is shown in Figure 8-8. A detailed computer program for estimating costs of



NOTES:
1. Minneapolis, Mar., 1972. ENR Construction Cost Index of 1827.
2. Amortization at 7% for 20 years.
3. Labor rate of $8.25 per hour.
4. Exhaust gas scrubber and enclosing structure included.
5. Costs do not include deodorization of gases: where required, add $4 to $10/dry ton.
6. Source: EPA Cost and Manpower Report and Stanley Consultants.

FIGURE 8-8. Multiple hearth incineration costs [6].

multiple hearth incineration has been developed. This program takes into account a large number of variables and is based on field data from nine operating municipal sludge incineration plants [7]. Some general cost curves for various plants have also been developed for multiple hearth furnaces [8]. The reported costs of multiple hearth sludge incineration vary substantially. The factors which cause these variations are the variability from one locale to another in the moisture content and nature of the sludge, size of the plant, chemicals used in dewatering, skill of operating and maintenance, labor and power, and fuel costs. For example, the Cleveland Southerly plant reports [9] power costs of $7.80 per ton of dry solids for vacuum filtration and incineration of 73,000 tons per year (about one-third of the total appears to be related to incineration). Rochester, New York, reports [10] an operating cost of $24.55 per ton for incineration of 390 tons of dry solids per year with the cost expected to drop to $15.40 per ton when the load increases to 2,180 tons per year. The South Lake Tahoe multiple hearth incineration system, which has a rated capacity of about 3,900 tons per year, reports an operating cost of $12.71 per ton [11].

8.2.5 Fluidized Bed Incineration

The first fluidized bed wastewater sludge incinerator was installed in 1962, and there are now several units operating in the United States. They range in size from 220 to 5,000 lb/hr dry solids. Ducar and Levin [12] have given a detailed description of several installations. A typical section of a fluid bed reactor used for combustion of wastewater sludges is shown in Figure 8-9. The fluidized bed incinerator is a vertical cylindrical vessel with a grid in the lower section to support a sandbed. Dewatered sludge is injected above the grid and combustion air flows upward at a pressure of 3.5 to 5.0 psig and fluidizes the mixture of hot sand and sludge. Supplemental fuel can be supplied by burners above or below the grid. In essence, the reactor is a single chamber unit where both moisture evaporation and combustion occur at 1,400 to 1,500° F in either the dense or dilute phases of the sandbed. All the combustion gases pass through the 1,500° F combustion zone with residence times of several seconds. Ash is carried out the top with combustion exhaust and is removed by air pollution control devices.

Several factors are involved in determining air flow. Fluidizing and combustion air must be sufficient to expand the bed and provide proper density to prevent sludge flotation. Excess air blows sand and products of incomplete combustion into the flue gases and depletes stored heat energy. Minimum oxygen requirements must be met to assure complete oxidation of all volatile solids in the sludge cake. Minimum temperatures must be maintained to assure complete deodorization, while maximum temperature limits are required to protect the refractory, heat exchanger, and flue piping. The quantities of excess air are maintained at 20 to 25 percent to minimize its effect on fuel costs as was illustrated by Figure 8-4. The capability of the fluidized bed furnace to operate at lower excess air than typically experienced in a multiple hearth furnace is one factor accounting for the greater heat efficiency of the fluidized bed system. The intense and violent mixing of the solids and gases within the fluidized bed results in uniform conditions of

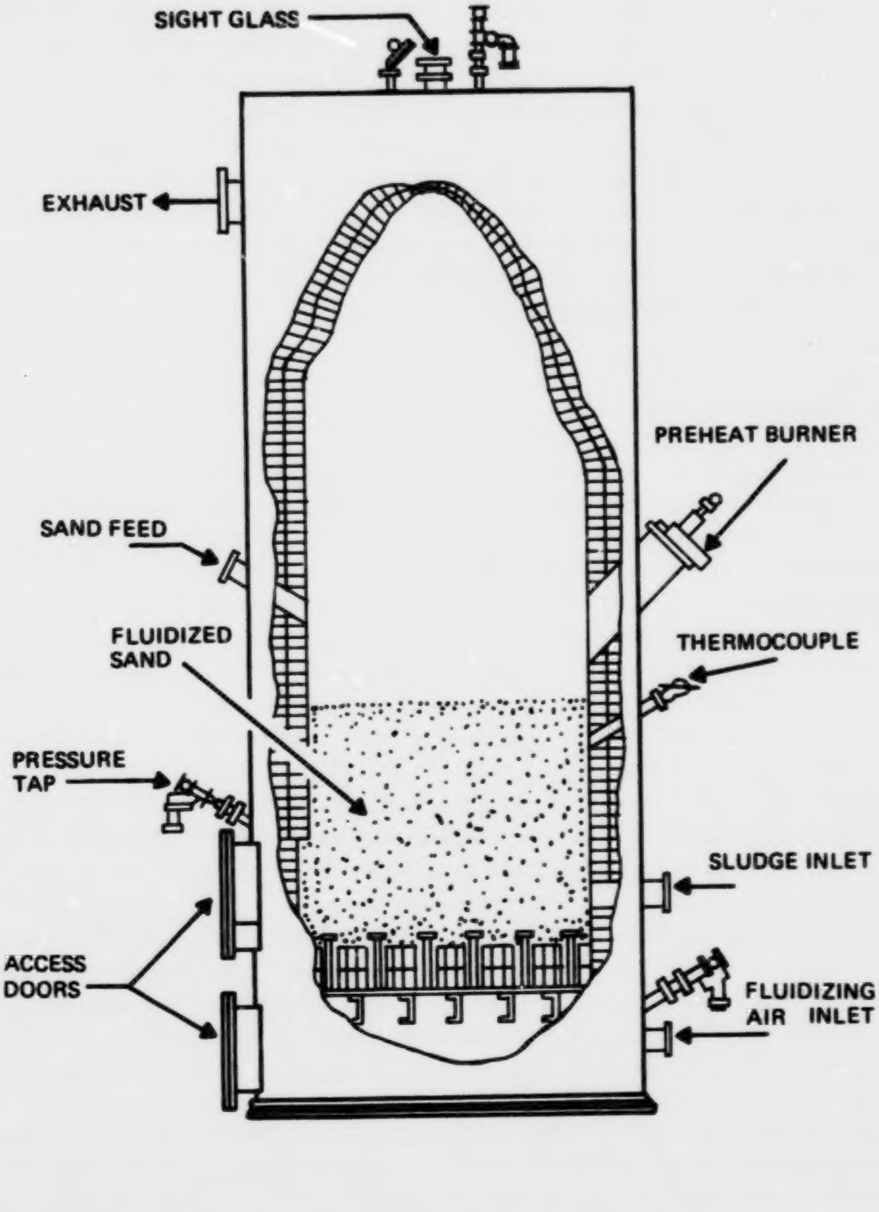

SIGHT GLASS
EXHAUST
PREHEAT BURNER
SAND FEED
FLUIDIZED SAND
THERMOCOUPLE
PRESSURE TAP
SLUDGE INLET
ACCESS DOORS
FLUIDIZING AIR INLET

FIGURE 8-9. Cross section of a fluid bed reactor.

temperature, composition, and particle size distribution throughout the bed. Heat transfer between the gases and the solids is extremely rapid because of the large surface area available.

As shown in Figure 8-10, an air preheater can be used in conjunction with a fluidized bed. Burd [13] reported that preheating the air from 70° to 1,000° F allowed a reduction in fuel costs from $9 to $3.50 per ton. Since air preheater cost can represent 15 percent of the fluidized bed furnace cost, a careful economic analysis is needed to determine its feasibility for a given situation.

The heat required for raising the sludge to the kindling point must come from the combustion zone. While standard combustion units rely on the heat transfer from the hot gases which contain only 16 BTU/cu ft, the expanded bed of the fluid-bed reactor has 16,000 BTU/cu ft. Because of the enormous reservoir of heat in the bed and a rapid distribution of fuel and sludge throughout the bed, optimum contact between fuel and oxygen and rapid transfer of heat is insured. The sandbed retains the organic particles until they are reduced to mineral ash, and the violent motion of the bed comminutes the ash material and prevents the buildup of clinkers. The resulting fine ash is constantly stripped from the bed by the upflowing gases. The heat reservoir provided by the sandbed also enables reduced start-up times when the unit is shut down for relatively short periods (overnight). As an example, a unit can be operated 4 to 8 hours a day with little reheating when restarting, because the sandbed serves as a heat reservoir.

Exhaust gases are usually scrubbed with treatment plant effluent and ash solids are separated from the liquid in a hydrocyclone, with the liquid stream returned to the head of the plant. An oxygen analyzer in the stack controls the air flow into the reactor and the auxiliary fuel feed rate is controlled by a temperature recorder. Shutdown controls for emergency situations should be provided.

The city of Lynnwood, Washington, installed the first commercial fluid bed system, and raw primary sludge is combusted following gravity thickening and centrifuge dewatering. Lynnwood's reactor is designed to receive 200 pounds per hour of dry solids with a moisture content of 65 percent and a volatility of 75 percent. The reactor has been operated with 20 percent excess air or about 360 scfm at a sludge feed rate of about 210 lb/hr dried solids. No. 2 oil is used for daily reheating and as auxiliary fuel, because the reactor has not been operated continuously. The reheat time and fuel required for reheating are a function of the duration of shutdown. The reheat time, Tuesday through Friday, is about 20 minutes, while for Monday it is about 1 hour [14].

The East Cliff Sanitary District, California, plant burns from 250 to 500 lb/hr of raw primary sludge in a residential area, and little auxiliary fuel is required. Auxiliary fuel is only required for start-up on Monday morning, because the bed temperature at shutdown exceeded 1,400° F and the overnight loss of temperature was less than 150° F. Reheating can be accomplished at a rate of 350 to 400° F per hour using 0.24 gpm of No. 2 oil. It has been found that the depth of the sandbed is not critical, and makeup sand is added once every one to two months. Maximum pressure within the unit is about 3.5 psig [14].

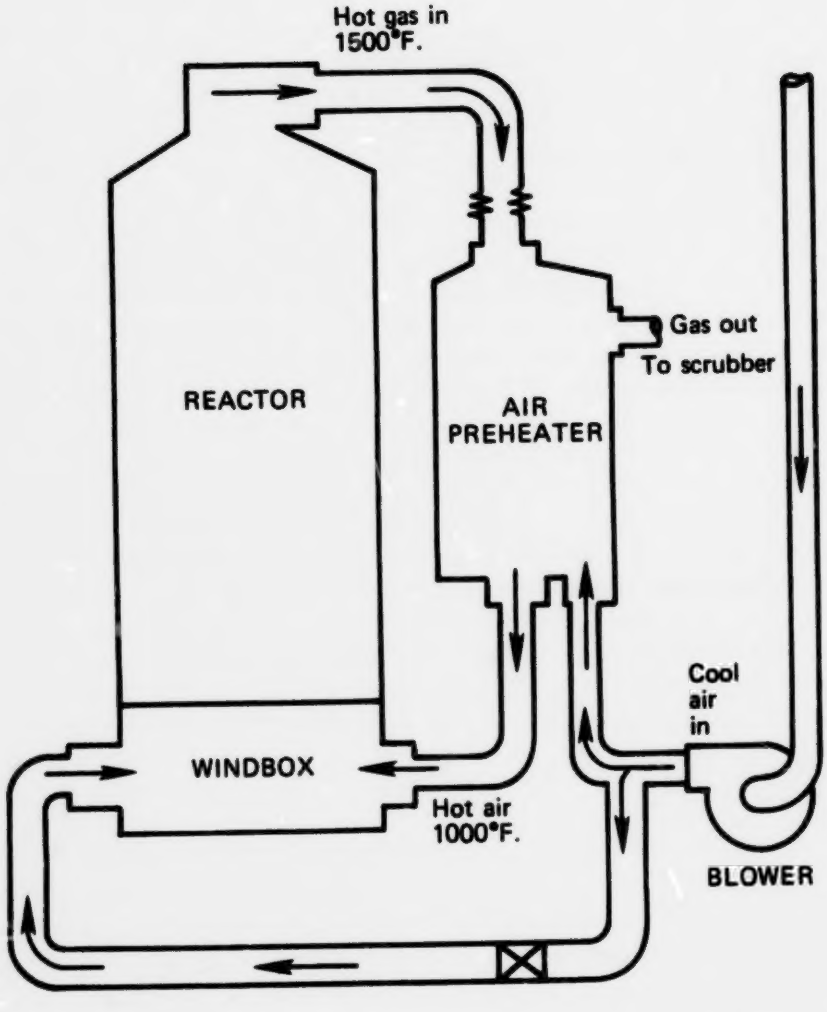

Hot gas in
1500°F.
Gas out
To scrubber
REACTOR
AIR
PREHEATER
Cool
air
in
WINDBOX
Hot air
1000°F.
BLOWER

FIGURE 8-10. Fluidized bed system with air preheater.

Alford [15] reported on the fluidized bed system at North Little Rock, Arkansas, and illustrated well the considerations typically involved in the operation of such a system. A centrifuged and degritted primary sludge with a solids concentration of 30 to 35 percent and a volatility of 78 percent was fed to a fluidized bed equipped with a preheater at a rate of 1,000 lb/hr of dry solids. The low-pressure high-temperature reactor has a diameter of 11.5 ft and an overall height of 28 ft. The bed was charged with 5,000 lb of high silica sand with an effective size of 0.210 mm and uniformity coefficient of 1.80. About 100 lb/month of sand is lost from the reactor by abrasion and disintegration. The hot sandbed is fluidized with outside air circulated through a single pass concentric core heat exchanger. This unit utilizes stack gas energy to heat the incoming air to about 950° F during normal operation. Shutdown of the unit is accomplished by first stopping the sludge feed system. Combustion air introduction is continued until the exhaust stack gases show about 15 percent oxygen, indicating near complete combustion of sludge in the bed. The fluidizing air feed is then shut off to retain stored heat energy. The temperature die-off curve normally is exponential and averages a rate of 8° F/hr from shutdown to restart. In the event the temperature in the reactor is still above 1,100° F by the time additional sludge is accumulated in the thickener, no auxiliary fuel is required for restart. Normally, the combustion process runs 15 to 18 hr/wk depending on the amount of sludge received.

Some extensive maintenance problems have occurred with preheaters used in fluidized bed systems, and scaling of the sand media has also been a problem. Most operating problems uncovered in a survey of many existing installations, however, were caused by jamming feed systems [12]. Screw feeds and screw pump feeds were both subject to jamming because of either overdrying of the sludge feed at the incinerator or because of silt carried into the feed system with the sludge. Another frequent problem has been the burnout of spray nozzles or thermocouples in the bed.

There is very little published data on the operation and maintenance (O/M) costs of fluid bed incineration. Albertson [16] has reported costs for fluidized systems operating mainly with primary sludge. The principal deterrent to development of field O/M costs has been the fact that most previous installations have been at smaller plants where record keeping has not been sufficient. One study collected capital cost data for several size systems [12]. In terms of February, 1968, dollars, the average capital cost of a 500 lb/hr dry solids system including the fluidized bed unit and centrifuge but excluding the preheater, buildings, installation, or engineering costs was $195,000. A 5,000 lb/hr system, on the same basis had a cost of $706,000 while a 250 lb/hr system had a cost of $122,000. One comparison [17] of the costs of multiple hearth and fluidized bed systems provides some insight into the inherent differences between the two systems. Capital costs for the two competing incineration systems are often essentially equivalent as demonstrated by results of bidding [18].

8.2.6 Flash Drying

Flash drying is the instantaneous removal of moisture from solids by introducing them into a hot gas stream. This process was first applied to the drying of wastewater sludge at the Chicago Sanitary District in 1932. The pictorial flow diagram of the C-E Raymond Flash Drying and Incineration System is shown in Figure 8-11 [19]. Originally, units were designed to dry sludge for fertilizer and burn only the excess.

The flash drying system shown in Figure 8-11 is based on four distinct cycles which can be combined in different arrangements. The first cycle is the flash drying cycle, consisting of the hot gas duct, cage mill, mixer, uptake duct, cyclone, air lock, dry divider, and vapor fan. The wet filter cake is blended with some previously dried sludge in a mixer to improve pneumatic conveyance. The blended sludge and the hot gases from the furnace at 1,300° F are mixed ahead of the cage mill and flashing of the water vapor begins. The cage mill mechanically agitates the mixture of sludge and gas and the drying is virtually complete by the time the sludge leaves the cage mill. The sludge, at this stage, is dry at a moisture content of 8 to 10 percent and the dry sludge is separated from the spent drying gases in a cyclone. The dried sludge can be sent either to fertilizer storage or to the furnace for incineration.

The second cycle is the incineration cycle. Combustion of fuel is essential to provide heat for drying the sludge and the fuel may be gas, oil, coal, or wastewater sludge. Primary combustion air, provided by the combustion air fan, is preheated and introduced at a high velocity to promote complete sludge combustion. The sludge ash accumulates in the furnace bottom and is removed periodically by a hydraulic sluicing system to an ash lagoon or other disposal area.

The third cycle is the effluent gas cycle or induced draft cycle consisting of the deodorizing and combustion air preheaters, dust collector, induced draft fan, and stack. Heat recovery is practiced to improve economy. The effluent gases then pass through a dust collector (dry centrifuge or wet scrubber) and the induced fan discharges the effluent gases through a stack into the atmosphere. The fourth cycle may be the fertilizer-handling cycle.

Perhaps the most notable current United States usage of this process is that by the city of Houston, Texas [19] primarily for drying sludge for use as a fertilizer. The total production capacity using the CE Raymond process was scheduled to reach 150 tons per day in 1973. In the present facilities at the Northside plant in Houston, excess activated sludge is dewatered on eight rotary vacuum filters and transported by belt conveyors to the flash drying system. It includes mixing vacuum filtered sludge with previously dried materials in a double paddle mixer before the combined solids are introduced into a cage mill into which hot gas from a furnace is injected. The dried product is discharged to a cyclone for separation of the solids from the entraining gases. The solids are fed partly to the product conveyor and partly to the double paddle mixer to help dry the incoming

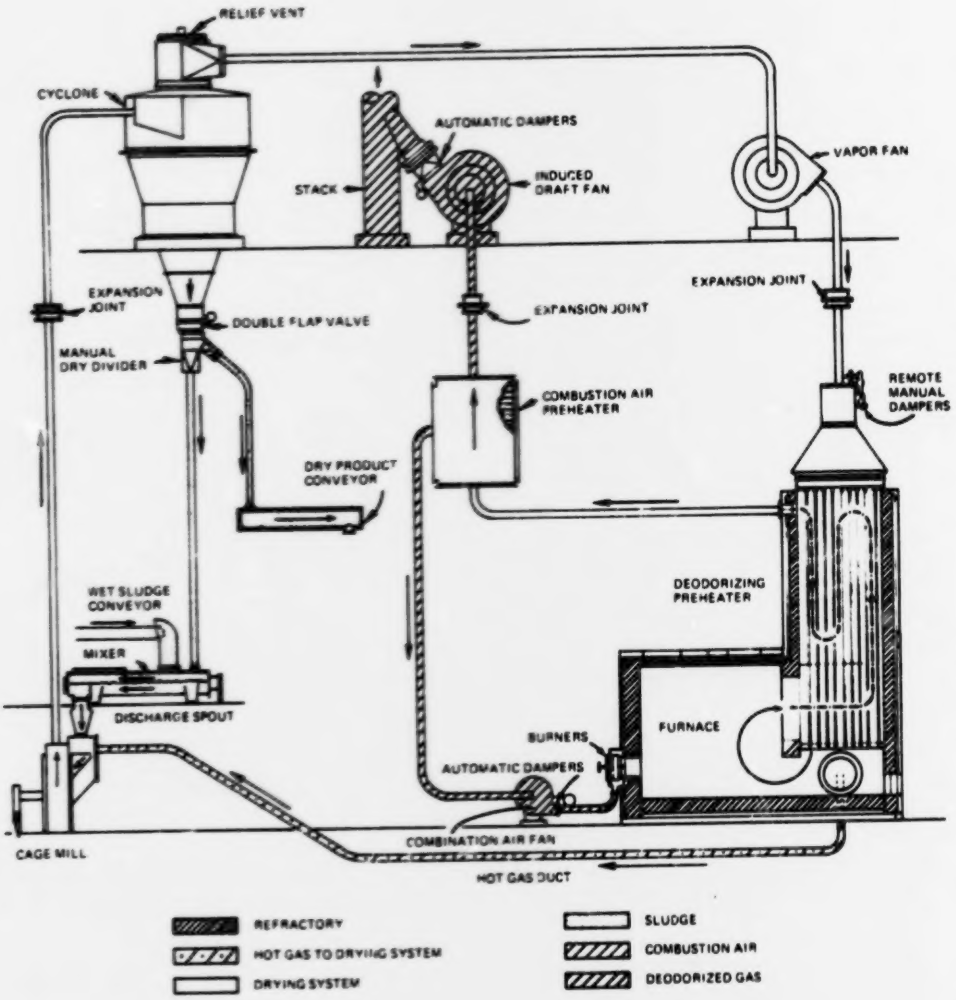

RELIEF VENT
CYCLONE
AUTOMATIC DAMPERS
VAPOR FAN
STACK
INDUCED DRAFT FAN
EXPANSION JOINT
EXPANSION JOINT
EXPANSION JOINT
DOUBLE FLAP VALVE
MANUAL DRY DIVIDER
COMBUSTION AIR PREHEATER
REMOTE MANUAL DAMPERS
DRY PRODUCT CONVEYOR
DEODORIZING PREHEATER
WET SLUDGE CONVEYOR
MIXER
DISCHARGE SPOUT
FURNACE
BURNERS
AUTOMATIC DAMPERS
CAGE MILL
COMBINATION AIR FAN
HOT GAS DUCT
REFRACTORY
HOT GAS TO DRYING SYSTEM
DRYING SYSTEM
SLUDGE
COMBUSTION AIR
DEODORIZED GAS

FIGURE 8-11. Flash dryer system [19].

sludge. The gases from the cyclone are recycled to the furnace where they pass through the deodorizer preheater section and are subjected to a temperature of 1,200° F for complete deodorization. The deodorized gases give up a portion of their latent heat to incoming vapor laden gases, and subsequently give up more heat to combustion air.

The temperature at the entrance to the cage mill where the hot gases and sludge to be dried are mixed is in the neighborhood of 1,200° F for evaporation of the moisture. At the cyclone, the temperature drops to around 220°F. In the subsequent heat exchange operation, the temperature of the effluent gases is reduced to about 500° F.

The process is automated and panel boards are provided in air conditioned cubicles which indicate and record variables such as air flow, temperatures at critical points, and amperage on fan motors. Horn alarms indicate unsuitable temperature conditions. The dried product is conveyed to a storage area for shipment. The organic constituents of the finished fertilizer product are about the same as in the mixed liquor and the principal variation is in the moisture content. The sludge on the filters averages about 3.4 percent solids and ranges from 2.8 to 4.0 percent. The dry product after complete processing has a moisture content of around 5.5 percent. From analysis at the time of sales of fertilizer in January, 1972, the moisture content was 5.0 percent; ash 34.76 percent; nitrogen 5.34 percent; and available phosphoric acid, 3.93 percent. The ash content fluctuates; the lowest on record is 26.4 percent and highest, 44.3. The variation apparently corresponds with drought and rainfall periods. The only artificial chemical conditioning used at the present time is to add ferric chloride to the sludge prior to filtration. In the usual operation, this amounts to about 75 lb per ton of dry solids or about 3.8 percent.

Throughout the experience with this operation, the city's marketing arrangements have been scheduled on the basis of competitive bidding. The successful bidder is committed to placing orders with the city for its entire production. At present, the contract period is five years, which is renewable. The revenue to the city in the first six months of 1972 averaged $21 per ton f.o.b. Houston.

The material is now shipped in bulk by railroad car lots or sometimes by barge. It is bagged for resale at the point of arrival. The present contractor has been handling it for about 10 years, disposing of about 80 percent of the production in the citrus groves of Florida. There has never been a time when it was not possible to dispose of the entire sludge production by sales.

Another approach to drying of wastewater sludges for use as fertilizers is currently being evaluated at the Blue Plains plant in Washington, D.C. [20]. A schematic of the system is shown in Figure 8-12. Drying is achieved in a jet mill in this case. The mill has no moving parts and offers the ability to dry and classify solids simultaneously.

As noted earlier, the use of the flash drying systems of the type shown in Figure 8-11 for incineration alone has not proven attractive. The Metropolitan Denver Sewage Disposal

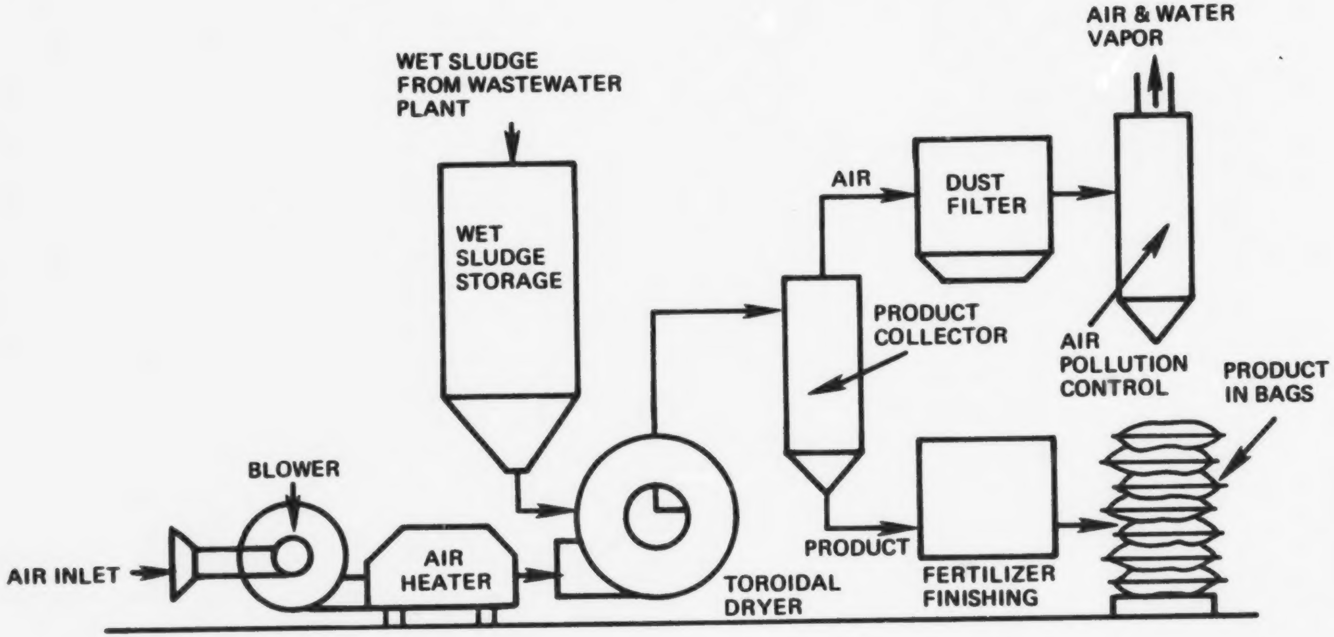

AIR & WATER VAPOR
WET SLUDGE FROM WASTEWATER PLANT
AIR
DUST FILTER
WET SLUDGE STORAGE
PRODUCT COLLECTOR
AIR POLLUTION CONTROL
PRODUCT IN BAGS
BLOWER
AIR INLET
AIR HEATER
PRODUCT
TOROIDAL DRYER
FERTILIZER FINISHING

FIGURE 8-12. Sludge drying system using the jet mill principle [20].

District No. 1 plant (approximately 100 mgd in capacity) abandoned a system of this type due to air pollution and problems of continuing explosions in the units. As an incineration unit, the flash drying system has the disadvantages of complexity, potential for explosions, and potential for air pollution by fine particles. An advantage is the flexibility it offers for drying a portion of the sludge for fertilizer.

Flash drying is relatively expensive because of fuel costs (contrasted to incineration–no heating value is realized from the sludge) and because pretreatment needs for production of sludge, which must have some reasonable nutrient balance, are also expensive. It has been reported that the fuel consumption for production of dried sludge is 8,000 BTU/lb.

Chicago, Illinois, reported a net cost of $45/dry ton for flash drying and abandoned the process because of cost and air pollution considerations. However, if recent increases in the values of fertilizers persist, the costs of sludge drying will be more largely recoverable from sale of the dried product.

8.2.7 Wet Air Oxidation

The wet air oxidation process is based on the principle that any substance capable of burning can be oxidized in the presence of liquid water at temperatures between 250° F and 700° F. The process can operate on difficult to dewater waste liquors and sludges where the solids are but a few percent of the water streams. In general, given the proper temperature, pressure, reaction time, and sufficient compressed air or oxygen, any degree of oxidation desired can be accomplished. By operating at lower temperatures and pressures, the same approach may be used for sludge conditioning as covered in Chapter 6.

The wet air oxidation process has been commercialized and patented as the Zimpro process [21]. This process has also been known as wet incineration, wet combustion, and wet oxidation processes. Wet air oxidation does not require preliminary dewatering or drying as required by conventional combustion processes. Water can be present up to 99 percent in this process whereas in conventional combustion it must be reduced to much lower levels to make incineration practical.

Another significant difference is the flameless oxidation of the organics at low temperatures of 300° F to 400° F when compared to 1,500° F to 2,700° F in the conventional combustion processes. Air pollution is minimized because the oxidation takes place in water at low temperatures and no flyash, dust, sulfur dioxide, or nitrogen oxides are formed.

The general flow diagram of the Zimpro continuous wet air oxidation system is shown in Figure 8-13. In the continuous process, the sludge is passed through a grinder which reduces any particles greater than ¼ inch to about ¼ inch in size. Sludge and air are then pumped into the system and the mixture is passed through heat exchangers and brought

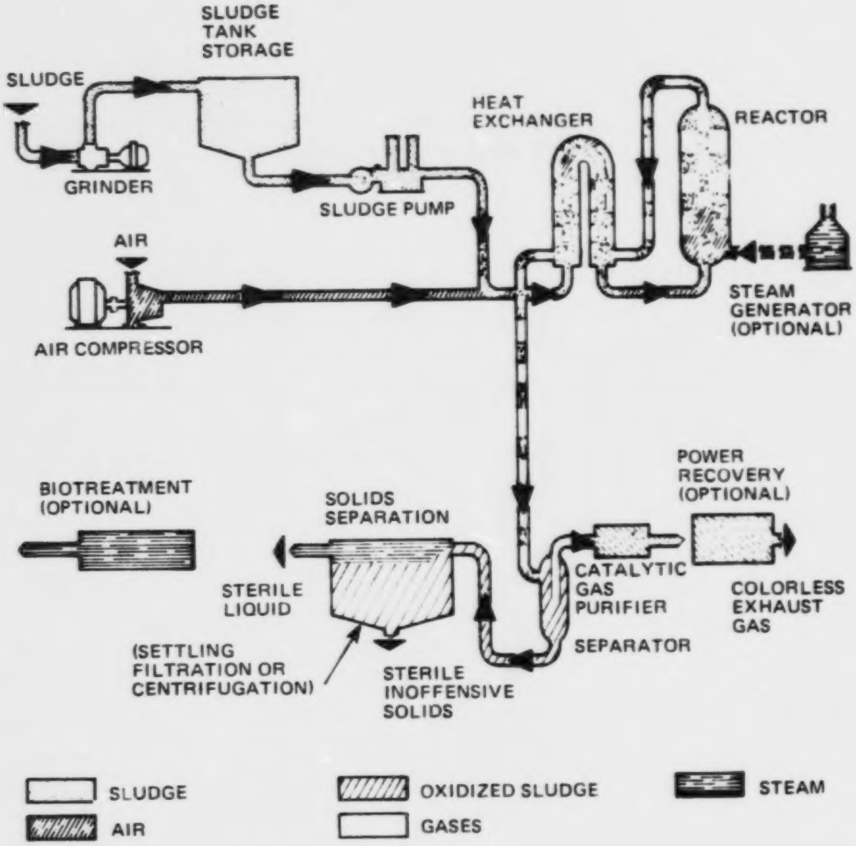

SLUDGE
SLUDGE TANK STORAGE
GRINDER
SLUDGE PUMP
HEAT EXCHANGER
REACTOR
AIR
AIR COMPRESSOR
STEAM GENERATOR (OPTIONAL)
POWER RECOVERY (OPTIONAL)
BIOTREATMENT (OPTIONAL)
SOLIDS SEPARATION
STERILE LIQUID
CATALYTIC GAS PURIFIER
COLORLESS EXHAUST GAS
SEPARATOR
(SETTLING FILTRATION OR CENTRIFUGATION)
STERILE INOFFENSIVE SOLIDS
SLUDGE
AIR
OXIDIZED SLUDGE
GASES
STEAM

FIGURE 8-13. Wet air oxidation system.

to the initiating reaction temperature. As oxidation takes place in the reactor, the temperature increases. The oxidized products leaving the reactor are cooled in the heat exchangers against the entering cold sludge and air. The gases are separated from the liquid carrying the residual oxidized solids and released through a pressure control valve to a catalytic oxidation unit for odor control. Where economic conditions make it attractive, the gases may be expanded in power recovery equipment before being discharged. The oxidized liquid and remaining suspended solids are released through a level control value and the solids may be separated by settling and drainage in lagoons or beds, or other methods such as vacuum filtration or centrifugation.

For start-up, heat is obtained from an outside source, usually a small steam generator. With high degree oxidations and high fuel value sludges, no external heat is needed once the process is started. Whenever the process is not thermally self-sustaining, steam may be injected continuously to sustain the reaction temperature.

Four important parameters control the performance of wet oxidation units: temperature, air supply, pressure, and feed solids concentration. The degree and rate of sludge solids oxidation are significantly influenced by the reactor temperature. A much higher degree of oxidation and shorter reaction times are possible with increased temperatures. As is the case in conventional incinerators, an external supply of oxygen (air) is required to attain nearly complete oxidation. The air requirement for the wet oxidation process is determined by the heat value of the sludge being oxidized, and by the degree of oxidation accomplished. Thermal efficiency and process economy are a function of air input, so it is important that the optimum amount be determined. Because the input air becomes saturated with steam from contact with reactor water, it is important to control the air also to prevent excessive evaporation of the water. For primary wastewater sludges with a BTU value of 7,800 BTU/lb, an air utilization of 5.75 lb/lb is typical. For an activated sludge with a heat value of 6,540 BTU/lb, an air utilization of 5.14 lb/lb would result.

Sufficient pressure must be provided to prevent water vaporization because operating temperatures are well above 212° F (typically above 500° F). Operating pressures have varied from 150 to 3,000 psi depending upon the degree of oxidation desired. Pressures are typically 1,000 to 1,750 psi in sludge handling installations.

The feed solids concentration has a significant effect on operating costs. It was found that if the solids concentration could be increased from 3 percent to 6 percent, the operating costs at the Chicago Sanitary District facility would decrease from $38/ton to about $23/ton [22]. The solids concentration is an important factor in keeping the oxidation self-sustaining.

A notable installation in the United States was the Chicago Sanitary District installation at its West-Southwest plant which was operated from 1962 to 1972 but has now been replaced by a land disposal system. This activated sludge plant treats over a billion gallons per day of wastewater and generates 800 tons of waste sludge every day. The 1962 installation was the first Zimpro plant erected in the U.S. for sludge treatment and disposal. Prior to installation of the wet oxidation system, of the 800 tons per day of sludge, up to 450 TPD was dewatered, heat dried, and sold as soil conditioner. Another 150 TPD of primary sludge was digested in Imhoff tanks and dried on sandbeds. The remainder went to lagoons. A pilot plant of the wet air oxidation process having a capacity of 2 TPD of sludge solids was set up and operated for about one year during 1957 to 1958. This work was successful and in 1959 a contract was signed calling for the erection of four units each having a capacity of 50 TPD of sludge solids at 3 percent solids concentration.

These units were placed in operation in 1962. It was soon evident that the plant had been very conservatively designed in terms of reactor capacity and amount of air supplied, and that the sludge processing capacity could far exceed the design value if thicker sludge could be obtained, or if the sludge pumping and heat exchange capacity could be increased. Because there was no means at hand for economically obtaining very much greater than 3 percent solids, it was decided to increase the sludge pumping and heat exchange capacity by approximately 40 percent. This expansion of the plant capacity was carried out while the plant was in operation by modifying one unit at a time. The nominal capacity of each of the modified units was 75 TPD with sludge at 3.4 percent solids, having 35 g/l COD. Usually the sludge concentration did not reach 3.4 percent and solids processing capacity was, therefore, proportionately less. The units accomplished approximately 70 percent COD reduction at the design capacity. At reduced solids throughput rates, COD reductions exceeding 80 percent were obtained.

The wet air oxidation plant routinely treated a mixture of raw primary and activated sludge solids in about equal weight proportions. The sludge was thickened by gravity settling to approximately 3 percent solids and then pumped through grinders to storage tanks in the oxidation facility. The sludge was ground to pass through 9/16 inch openings. Approximately 0.3 g/l of ammonia was added to the sludge on its way to the storage tanks to adjust the pH to approximately 7.0, which minimized scaling in heat exchangers. Sludge was taken from the storage tanks by centrifugal pumps that increased the sludge pressure to about 50 psig. These fed the positive displacement high pressure pumps which raised the sludge pressure to about 1,800 psig. Compressed air was then introduced into the sludge at the same pressure and the mixture was passed through heat exchangers. The sludge-air mixture was brought to approximately 400 to 420° F and introduced into the reactor where the oxidation occurred. The oxidized material, consisting of water, ash, steam and noncondensible gas, left the reactor at 500 to 520° F and passed back through the heat exchangers into the separator. In the separator the water carrying the ash was separated from the stream and noncondensible gases. The gas and steam passed either to the turbine for power generation or directly to a water scrubber from which they were exhausted to the atmosphere. The oxidized sludge was discharged through a coil in the storage tank for further cooling and eventually was discharged to a lagoon, where the ash was settled and the supernatant was returned to the treatment plant. Tests indicated that vacuum filtration at 5.6 psf/hr would produce a cake with 40.5 percent solids and with filtrate solids of 0.16 to 0.6 percent with no chemical conditioning. Centrifuging also appeared promising. Solids captures of 60 to 70 percent could be obtained, starting with a 10.8 percent solids feed, by centrifuging. The dewatered ash contained 52 to 56 percent solids. Solids captures greater than 99 percent were possible with the use of reasonable doses of an anionic polyelectrolyte; however, the cake solids concentration dropped to 30 to 40 percent [23].

Each oxidation unit was normally on line for a period of 30 to 60 days. When heat exchange efficiency was noticeably reduced, the unit was shut down, the heat exchangers isolated from the rest of the system and washed by circulating acid or caustic through them. This removed any accumulated scale and the unit was returned to service in 24 to 48 hours. Air compressors and sludge pumps also required periodic maintenance. Reactors were emptied on an irregular basis, but no more than once yearly cleanings appeared to be necessary.

Following the expansion of the facility in 1967, Chicago processed an average of about 248 tons/day over the period 1967 to 1972 at an average cost of about $35/ton. However, late in 1972, the Chicago Sanitary District shut down all of the units in favor of land disposal of the organic sludges. The Chicago units have not been operated since late 1972. The compressors are operated weekly and the other mechanical components maintained to keep the wet air oxidation units available as backup to the land disposal systems. The Sanitary District reports an improved secondary effluent quality at the West-Southwest plant since the wet oxidation recycle liquors have been removed from the system.

Wet oxidation of raw primary sludge has been practiced at Rye, New York [24]. The Blind Brook treatment plant at Rye achieved a 90 percent reduction of insoluble organic matter by operating at a temperature of 237.8° C and a pressure of 750 psi. The wet oxidation facility has been operated intermittently on a 7 days on, 7 days off, schedule. Auxiliary fuel has been used only when starting the unit. The oxidized sludge (ash) had an organic content of 18.6 percent during the first year of operation (1964). After cooling and solid-liquid separation, the BOD of the supernatant effluent averaged 8,400 mg/l. Separated ash disposal has been a problem at times due to odors from the ash drying beds. In late 1966 [25] the operators reported few maintenance problems. Some occasional problems had occurred from blockage of heat exchange tubes and failure of a reactor baffle.

The operating costs for the plant at Rye, New York, have been reported as $26.80 per ton made up of the items below:

Power = $13.60/ton ($0.023/KWH)

Chemicals = 3.60/ton

Water = 3.60/ton

Labor = 6.00/ton

McKinley [26] described the Wheeling, West Virginia, wet oxidation capital and operating costs as:

1. Installed capital cost = $284,000 for a 5.6 ton/day facility.
2. Operating cost = $19.97/ton on which was broken down as follows:

Power = $6.11/ton

Chemicals = 4.13/ton

Fuel = 1.65/ton

Maintenance = 1.17/ton

Labor = 6.91/ton

An insoluble organic destruction of 90 percent was achieved, starting with a raw primary sludge feed of 7.35 percent solids.

The above cost estimates do not include the cost of handling the recycled liquors or the oxidized solids. At Chicago [23], the estimated cost of separating the oxidized solids was:

Operation	Estimated Cost/ton
Sedimentation Thickening	$0.27
Vacuum Filtration	0.30
Centrifugation	1.60

An operational disadvantage of the wet air oxidation technique is the need to recycle wet air oxidation liquors, high in organic content and in phosphorus and nitrogen, back through the wastewater treatment process. These liquors represent a considerable organic load on the treatment system and the costs of handling these liquors is not usually included in reported costs. The BOD contained in these liquors may be 40 to 50 percent of the BOD of the unprocessed sludge and is typically high in ammonia and volatile acids resulting from the oxidation of the nitrogen and carbon constituents of the sludge [27,28]. The BOD of recycled liquors is typically 4,000 to 5,000 mg/l. They also contain 400 to 1,000 mg/l NH_3-N and 7,000 to 10,000 mg/l COD. The pH is acid, in the 5 to 6 range usually.

Studies [29,30] of aerobic, biological treatment indicate that the liquors are amenable to biological treatment with BOD removals of 90 percent or greater achieved at loading rates of 4 lb BOD/lb of mixed liquor volatile solids. It was estimated that if the return liquors receive separate aerobic, biological treatment prior to recycle to the main biological plant that the plant effluent BOD will not be increased and the effluent COD will increase by 8 percent. Direct recirculation of untreated liquors was estimated to cause a 10 percent increase in effluent BOD and 20 percent increase in effluent COD–assuming the capacity is available in the main plant to handle the recycled load [30].

The COD noted above to be refractory to the biological processes has also been found difficult to remove by advanced waste treatment processes. A pilot study at Colorado Springs, Colorado, found that some of this recycled COD would also pass through coagulation and activated carbon adsorption processes downstream of the secondary process.

The treatment of recycle liquors prior to return to the main plant is prudent to minimize variations in the applied load to the main plant. Recognition of the magnitude of the recycled load and incorporation of the capacity to handle it is imperative to successful plant operation.

Among other design considerations are the facts that odor problems can develop from the off-gases and from lagooning of the ash containing effluent. Air pollution caused by the stack gases can be controlled by catalytic burning at high temperatures. Odors from lagooning or sand drying bed operations might best be solved by dewatering the ash in a system that includes gravity separation thickening followed by dewatering on vacuum filters or in centrifuges.

Frequent shutdowns and maintenance problems have been reported from several installations. The high pressure-high temperature system also introduces some safety hazards. Safety considerations must play an important role in plant design.

8.2.8 Pyrolysis

Like incineration, pyrolysis is a controlled combustion process. Unlike the term incineration, however, pyrolysis does not imply that a waste is being burned. The pyrolysis process has been used for years by industry–for example, production of charcoal and methanol from wood, and coal gasification. The process requires raising the fuel to a temperature at which the volatile matter will distill, leaving carbon and inert material behind. The carbon and volatiles do not burn in the process because the heating occurs in an atmosphere deficient in oxygen. Volatile matter may be burned off as waste in a secondary chamber to which air is added, or the off-gas may be cooled and condensed to recover oils and tars or cleaned and used as fuel. Like incineration, pyrolysis reduces the sludge volumes and sterilizes the end product. Unlike incineration, it offers the potential advantages of eliminating air pollution and producing useful by-products. Air pollution can be controlled because heating takes place in a closed system that allows the collection of gases for beneficial use as a fuel or controlled burning. Those systems ready for marketing are offered on a proprietary basis in which design, construction, and usually operation are offered on a turnkey basis. Cost data from full-scale operating systems are not yet available, and there is no basis on which to assess the reliability of these new plants. Available information [31,32], however, indicates that the developers of the new systems are willing to proceed at costs which are initially competitive with conventional incineration.

8.2.9 Other Types of Incinerators

Cyclonic

Cyclonic reactors are designed for sludge disposal in the smaller wastewater treatment plants. The principle of the cyclonic reactors is that high velocity air, preheated with combustion gases from a burner is introduced tangentially into a cylindrical combustion chamber. Concentrated sludge solids are sprayed radially towards the intensely heated walls of the combustion chamber. This feed is immediately caught up in the rapid cyclonic flow of hot gases and combustion takes place rapidly so that no material adheres to the walls. The ash residue is carried off in the cyclonic flow and passes out of the reactor.

Typically, degritted, thickened primary plus activated sludge is pumped from a thickener to a centrifuge. The centrifuge dewaters the cake, which drops into a hopper and is subsequently pumped into the cyclonic reactor with a small amount of compressed air.

One feature that makes this sludge combustion system attractive for some small communities is that the unit is often sold premounted on skids as a package system. Figure 8-14 shows a view of one type reactor with the sludge hopper mounted on the skid as it would be shipped to the job. The system dewatering components may also be skid mounted and the skids interconnected in the field.

The system lends itself to compact layouts. The dewatering and incinerating portions of the system for a plant serving 5,000 people occupies a space 18 ft × 18 ft. The unit also requires relatively short start-up times and may be brought to operating temperature (above 1,400° F) in less than one hour.

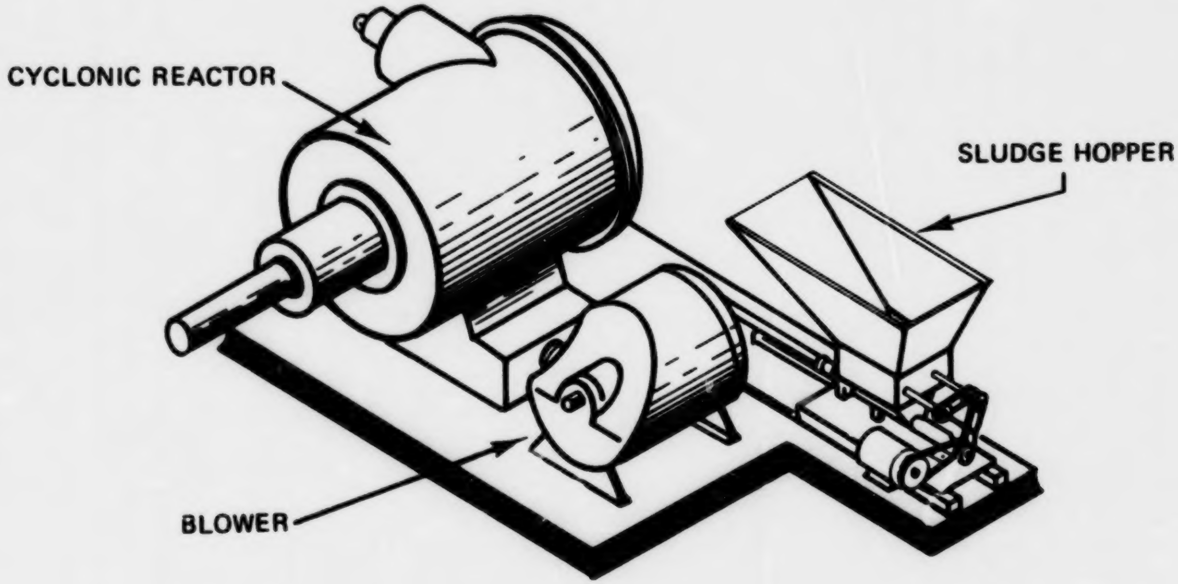

CYCLONIC REACTOR
SLUDGE HOPPER
BLOWER

FIGURE 8-14. Skid-mounted cyclonic reactor system [33].

These reactors process combined primary plus secondary sludge at nominal rates up to 100 to 130 pounds of dry solids per hour or 500 to 650 pounds of wet sludge per hour. The detention time for the sludge within the reactor is less than 10 seconds. The temperature is kept above 1,400° F so that the organic matter is burned above the odor producing level. The first installation is at the Richardson Bay Wastewater Treatment Plant at Tiburon, California.

Figure 8-15 shows another system using a cyclonic reactor applicable to larger plants is also now available [34]. The unit is designed as a vertical cylinder with a rotating, solid hearth on which the material for combustion lies. Unlike conventional forms of incinerators, no combustion air is allowed to pass upwards through the material. All of the air for combustion is injected into the furnace through tangential high velocity inlets which impart a cyclonic, swirling pattern to the movement of the gases which pass over the top of the burning material assuring an adequate contact of oxygen. The products of combustion enter a central vortex before leaving the furnace through a top conical or domed outlet. All of the solids enter the furnace at a minimum furnace temperature of 600 to 800° C with the initial temperature achieved by preheating with auxiliary fuel. Dewatered solids are stored and are fed to the furnace by a transfer screw conveyor which deposits the solids on the edge of the rotating hearth by means of a second water cooled conveyor. Once on the hearth, the material travels in a predetermined depth and width around the outer annulus as the hearth rotates, being subsequently ploughed over towards the center of the hearth after one revolution by a fixed plow. Subsequent revolutions of the hearth continue to move the material across towards the center in a series of concentric annular paths until finally the incombustible material is discharged through an ash chute in the center. This cyclone furnace is a British development and is available in capacities to 6 tons/hr. There are no current applications in the U.S. at this time.

Electric

An all electric furnace using an infrared heat source is currently under development by Shirco Co., Dallas, Texas, with the first full-scale sludge incineration units scheduled for Richardson, Texas (500 lb/hr) and Greenville, Texas (900 lb/hr). Recent developments in infrared lamps, coupled with the advent of silicon controlled rectifiers, semiconductor controls and ceramic reflector materials, have provided an economical means for applying and controlling radiant energy.

By properly applying this technology, it may be possible to approach operating costs of natural gas with an all electric infrared incinerator utilizing smaller gas scrubber systems, less combustion air, and lower capital investments than other incinerators presently on the market. The shortage of natural gas and oil supplies is a factor enhancing the appeal of this approach. Also, areas remote from natural gas or other petroleum fuel sources may find electric infrared incineration an attractive alternate for ultimate disposal of sludge.

Figure 8-16 is a schematic of the overall system. The dewatered sludge is conveyed to the belt conveyor which discharges the sludge into the machine onto a conveyor belt. The

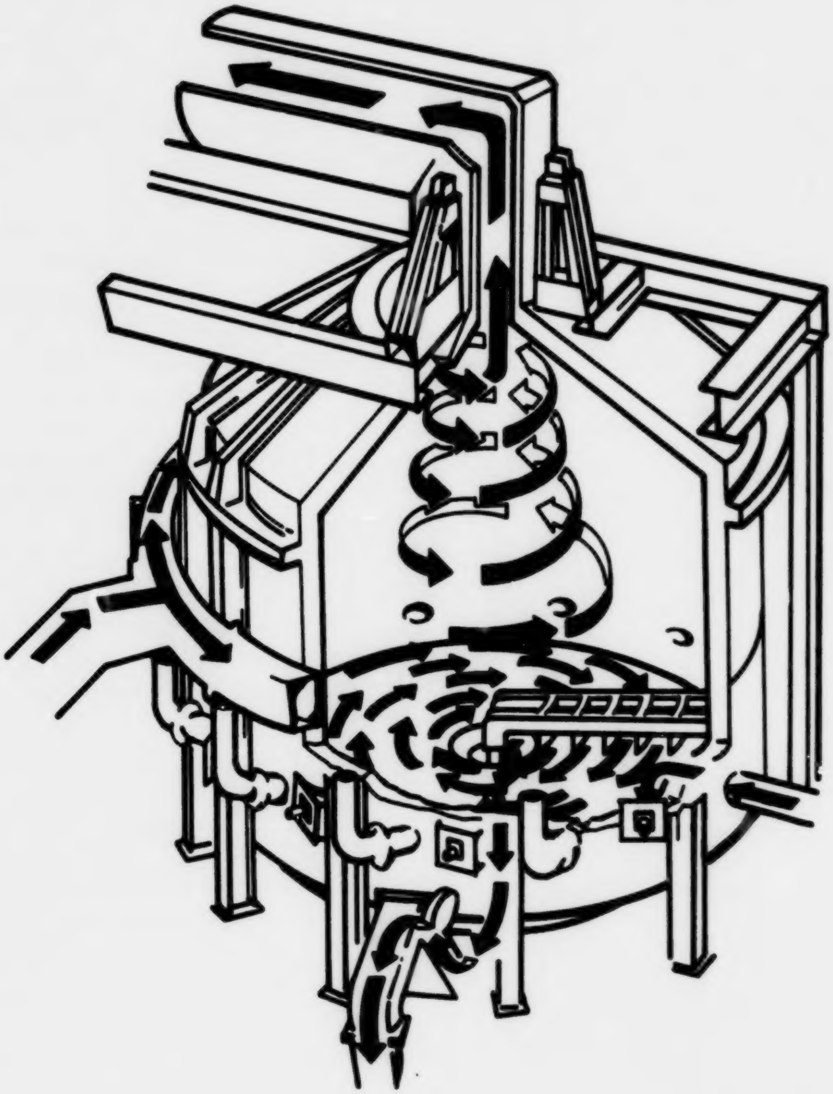


FIGURE 8-15. Cyclone furnace.

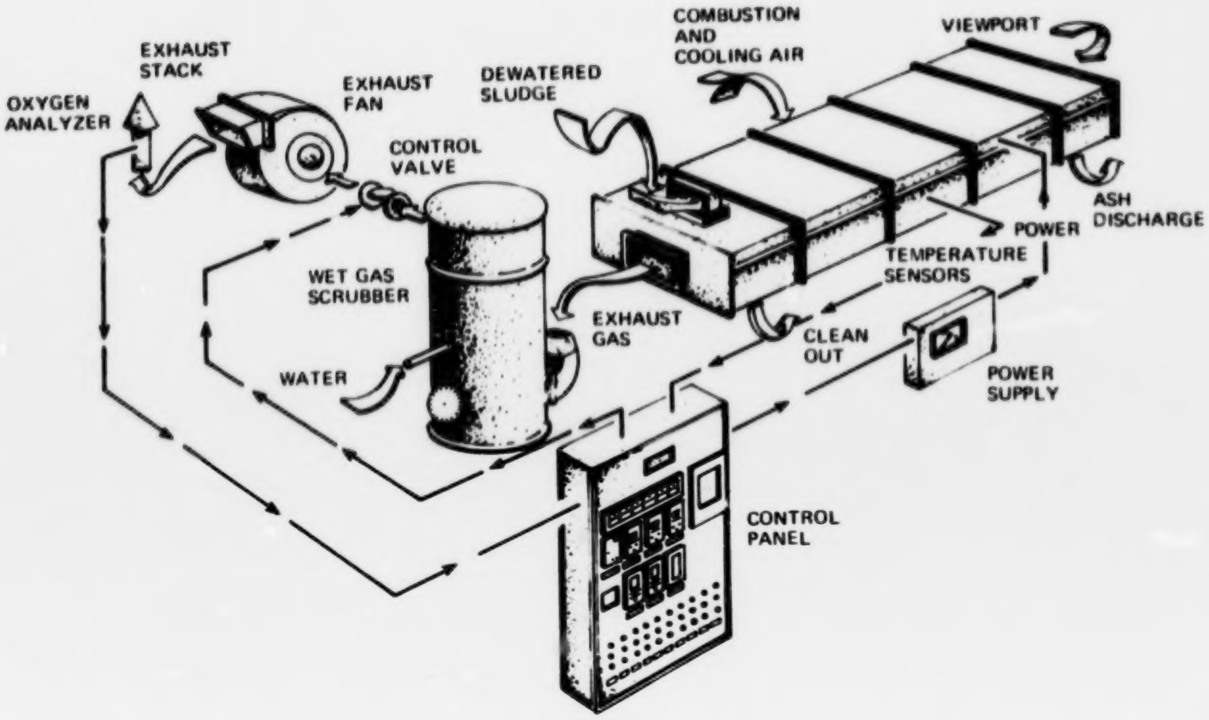

EXHAUST STACK
OXYGEN ANALYZER
EXHAUST FAN
CONTROL VALVE
DEWATERED SLUDGE
COMBUSTION AND COOLING AIR
VIEWPORT
ASH DISCHARGE
POWER
TEMPERATURE SENSORS
WET GAS SCRUBBER
EXHAUST GAS
CLEAN OUT
WATER
POWER SUPPLY
CONTROL PANEL

FIGURE 8-16. Infrared incineration system.

high temperature belt conveyor carries the sludge through a drying zone and then into a combustion zone. In the combustion zone, mounted just above the belt, is a battery of infrared lamps which initiates and maintains the combustion. The belt then discharges the ash into a hopper at the end of the machine. The lamps and end seals are cooled by drawing outside air through the cooling air ducts. This preheated air is then used as combustion air. The combustion air is then exhausted through a wet gas scrubber or necessary air pollution equipment.

Because the heat is transferred by radiation rather than conduction or convection, the air is not heated reducing combustion air requirements. The potential advantages of this system appear to be:

- Low capital cost.
- Low operation and maintenance costs.
- The incinerator can be brought from ambient temperature to 1,600° F to 1,800° F within one hour.
- There is no explosion danger.

Although there is no doubt that natural gas or oil provides a cheaper source of heat than electricity, the other savings associated with the infrared system may offset the higher unit fuel costs. The infrared approach also scales down much better than other systems and may be particularly attractive for small plants. This system also shows potential for the regeneration of activated carbon. A 50 lb/hr unit is in operation for carbon regeneration in an industrial application in Baton Rouge, Louisiana.

8.3 Lime Recalcining

Lime is often used as a coagulant either as a tertiary step or ahead of the primary clarifier in either a biological or a physical-chemical plant for removal of phosphorus from wastewaters. There is, of course, considerable experience around the world with the successful recalcining and reuse of lime used in water treatment plants and these techniques may also be used to recalcine and reuse lime in wastewater applications.

The process of recalcining consists of heating the dewatered calcium-containing sludge to about 1,850° F which drives off water and carbon dioxide leaving only the calcium oxide (or quicklime). When dealing with wastewaters, the lime sludge contains inert materials which must be wasted from the system or an infinite buildup in the quantity of sludge to be handled will occur. When coagulating secondary effluent, these inert materials are made up largely of magnesium hydroxide and hydroxyapatite. However, when coagulating raw wastewater, the lime sludges will contain the many inert solids found in raw wastewater and the magnitude of the inert problem increases significantly. As discussed in Chapter 7, centrifugation may be used to classify the inert and largely remove them from the lime sludge.

The only significant experience in the U.S. with recalcining lime sludges from the tertiary coagulation of municipal wastewaters is that gained from the South Lake Tahoe plant [11,35]. At the design flow of 7.5 mgd through this water reclamation plant, approximately 17 tons (dry CaO basis) per day of lime mud would have to be dewatered and disposed of. Since about 93 percent by weight of this lime mud is in the form of calcium carbonate, disposal costs would include not only dewatering and disposing of about 34 tons of water and solids but also the loss of recoverable calcium oxide. By recovering the lime through recalcination, the total blowdown of waste solids is reduced to about 1.5 tons of dry solids. The cost of recalcined lime is slightly more than that of new lime at this specific 7.5 mgd plant; however, the reuse of lime reduces by a factor of 20 the amount of water and sludge to be disposed of and, therefore, effects a substantial overall cost savings.

The South Lake Tahoe system is shown schematically in Figure 8-17. Lime mud is pumped from the chemical clarifier and recarbonation reaction basin to a gravity thickener. Thickened lime mud is then pumped to 24" X 60" solid bowl concurrent flow centrifuges operated in series for classification (see Chapter 7).

The cake from the first centrifuge is carried by a belt conveyor to a 14.3 foot diameter, six hearth furnace in which calcium oxide and carbon dioxide are produced. The recalcined lime is conveyed out of the furnace by gravity through a crusher to a thermal disc cooler where lime temperatures are lowered from 700° F to 100° to 150° F, and then into a rotary air lock. The recalcined lime is pneumatically conveyed from the rotary air lock to a 35-ton capacity recalcined lime storage bin for eventual reuse. Stack gases, rich in carbon dioxide, are scrubbed in a multiple tray scrubber before being exhausted to the atmosphere. A portion of the gases are recycled to the recarbonation system to adjust the pH of the lime coagulated wastewater to about 7.

Since April, 1968, the Lake Tahoe plant has successfully recalcined lime sludge from the lime chemical treatment process. Over this period makeup lime has accounted for 28 percent of the calcium oxide used. Average monthly CaO values in the recalcined lime have ranged between 51.0 percent and 74.7 percent with the average over the entire period being 66.0 percent. It has been found that about 3.7 percent by weight of the usuable calcium entering the furnace is lost as fly ash and captured by the wet scrubber.

At Tahoe, varying the temperatures between 1,600° F and 1,900° F has a major effect on recalcined lime activity. Within this temperature range there is no indication that the lime is being overburned. Recalcining lime at 1,900° F as opposed to 1,800° F produced a 5 percent increase in available calcium oxide, but very little improvement in an already acceptable slaking rate. At 1,600° F the flour-like recalcined lime showed pronounced

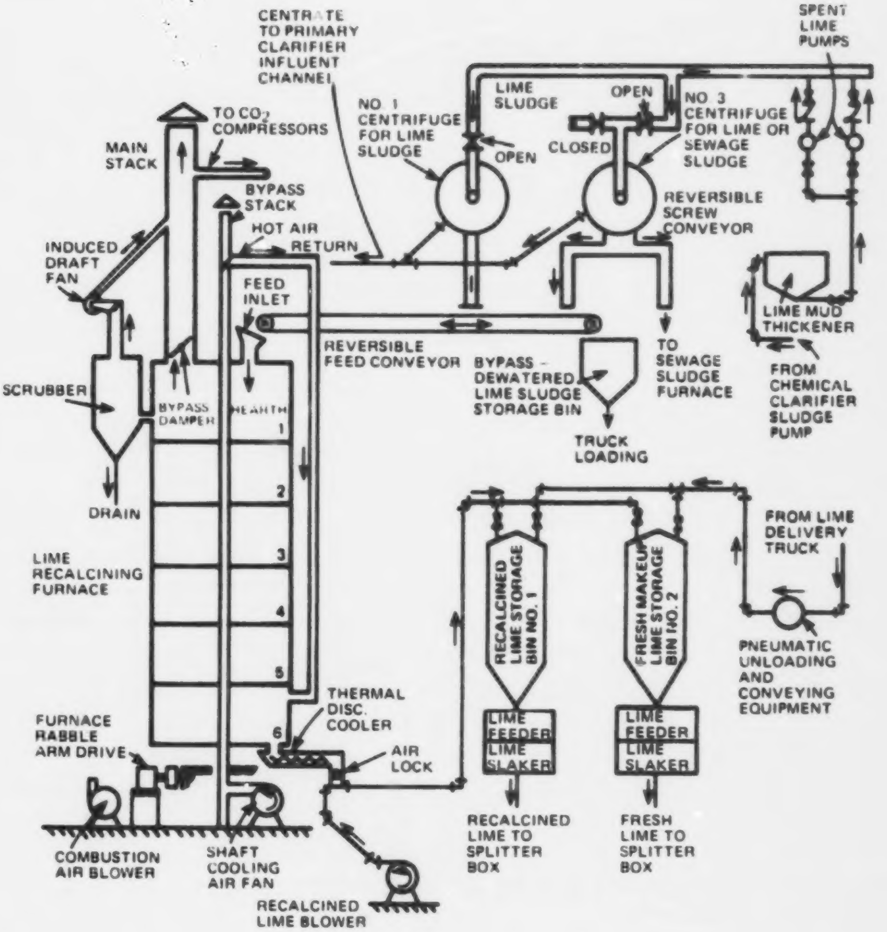

CENTRATE TO PRIMARY CLARIFIER INFLUENT CHANNEL
SPENT LIME PUMPS
TO CO2 COMPRESSORS
MAIN STACK
LIME SLUDGE
OPEN
NO. 1 CENTRIFUGE FOR LIME SLUDGE
OPEN
CLOSED
NO. 3 CENTRIFUGE FOR LIME OR SEWAGE SLUDGE
BYPASS STACK
REVERSIBLE SCREW CONVEYOR
HOT AIR RETURN
INDUCED DRAFT FAN
FEED INLET
LIME MUD THICKENER
REVERSIBLE FEED CONVEYOR
BYPASS - DEWATERED LIME SLUDGE STORAGE BIN
TO SEWAGE SLUDGE FURNACE
FROM CHEMICAL CLARIFIER SLUDGE PUMP
SCRUBBER
BYPASS DAMPER
HEARTH
1
TRUCK LOADING
2
DRAIN
FROM LIME DELIVERY TRUCK
3
LIME RECALCINING FURNACE
RECALCINED LIME STORAGE BIN NO. 1
FRESH MAKEUP LIME STORAGE BIN NO. 2
4
PNEUMATIC UNLOADING AND CONVEYING EQUIPMENT
5
THERMAL DISC. COOLER
LIME FEEDER
LIME SLAKER
LIME FEEDER
LIME SLAKER
FURNACE RABBLE ARM DRIVE
6
AIR LOCK
RECALCINED LIME TO SPLITTER BOX
FRESH LIME TO SPLITTER BOX
COMBUSTION AIR BLOWER
SHAFT COOLING AIR FAN
RECALCINED LIME BLOWER

FIGURE 8-17. The lime recalcining system at south Lake Tahoe.

tendencies to agglomerage into soft, easily crushed particles of ¼ to ¾ inch diameter. Many of the particles contained centers of unburned organic sludge.

The optimum furnace conditions in terms of recalcined lime activity appear to be about 1,900° F on the fourth and fifth hearths at 1.5 to 2.0 rpm rabble rate.

The large quantities of inert materials found in the sludges resulting from lime coagulation or raw wastewater make lime recovery and reuse more difficult. One plant is now under construction at Contra Costa, California, where a system for such lime recovery and reuse has been developed which uses a combination of wet and dry classification techniques [36,37]. At Contra Costa, a mixture of lime (to pH 11) and ferric chloride (about 14 mg/l) are added to the raw wastewater as coagulants. Following primary clarification, the wastewater receives biological treatment, including biological nitrogen removal.

The lime sludges are thickened and then passed through series centrifugation for classification in a manner similar to that described for the Tahoe project. Pilot data indicate that 90 percent calcium capture is possible in the first stage centrifuge while rejecting 40 to 85 percent of the inert materials. Following recalcination in a multiple hearth furnace, the ash discharge from the furnace will be passed through a dry classification device for further purging of inert materials. The dry classification device makes a separation with air based on particle size while the preceding centrifugal wet classification is based on particle weight. Based upon pilot tests, the use of these two classification techniques in series offers a means of providing adequate purging of inerts to permit lime reuse when coagulating raw wastewater.

Although the Lake Tahoe and Contra Costa waste treatment plants are both utilizing the multiple hearth furnaces, the fluidized bed furnaces discussed earlier may be used and have been used successfully for recalcining in water treatment plants for many years. The fluidized bed system for lime recalcining is shown in Figure 8-18 [38,39]. Lime mud filter cake is fed into a paddle mixer along with dry recycled fines and quench water. The mixture then goes to a cage mill disintegrator where precooled calciner stack gas at 1,000° F dries and disintegrates the moist solids. The resultant fine, carbonate is conveyed by the exhaust gas to a cyclone separator. Large fraction discharge from the cyclone is split, a portion being recycled to the mixer and a portion to the calciner feed bin.

Calcination takes place in a two-compartment fluidized bed furnace. The upper fluid bed of the reactor is used for low temperature calcination (1,500° to 1,600° F) of calcium carbonate and the pelletization of calcium oxide. The lower fluid bed cools the calcined product. In both fluid beds the solid particles are supported on a rising column of air so that the solids behave in much the same fashion as a liquid. Fresh solids added to the bed are quickly and uniformly distributed. The beds are held in a constant state of agitation and suspension so that heat transfer is instantaneous and uniform.

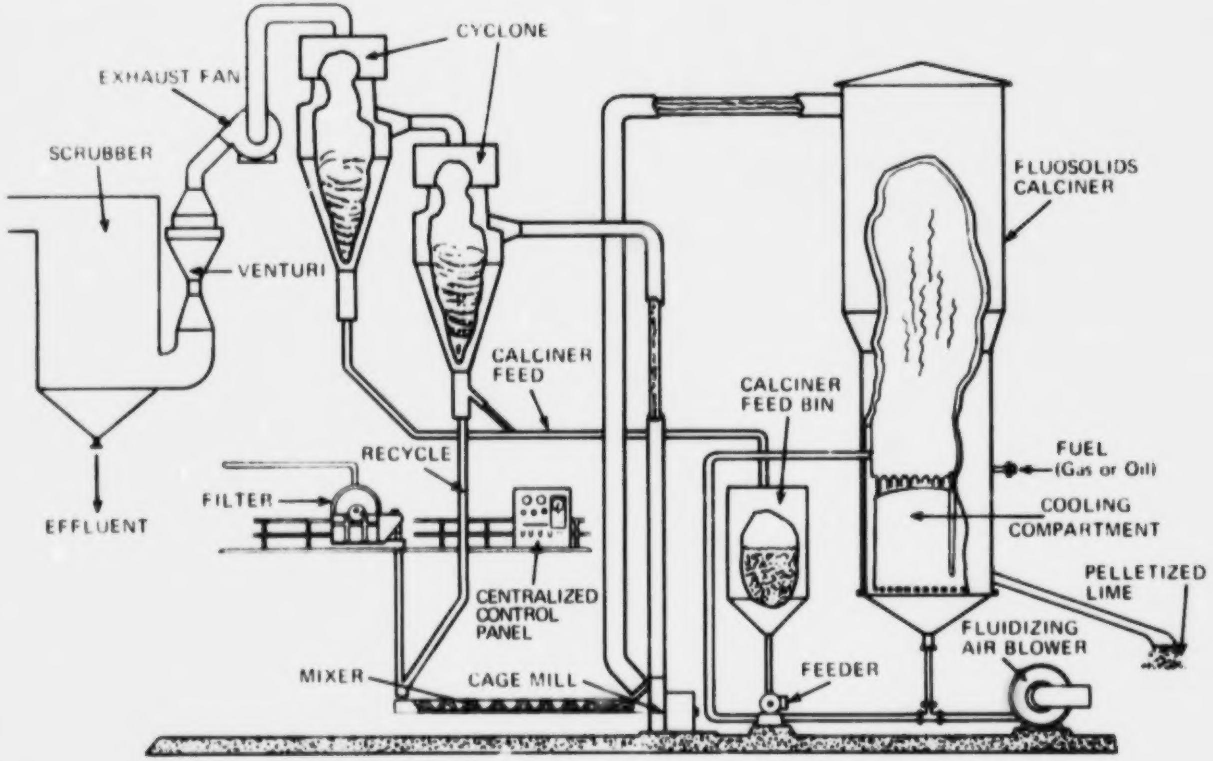

CYCLONE
EXHAUST FAN
SCRUBBER
VENTURI
EFFLUENT
FILTER
RECYCLE
CALCINER FEED
CENTRALIZED CONTROL PANEL
MIXER
CAGE MILL
CALCINER FEED BIN
FEEDER
FLUOSOLIDS CALCINER
FUEL (Gas or Oil)
COOLING COMPARTMENT
PELLETIZED LIME
FLUIDIZING AIR BLOWER

FIGURE 8-18. Fluidized bed system for lime recalcining.

Lime produced by this system is in the form of pelletized particles, 6 to 20 mesh in size. These uniform spheres are soft-burned, dust-free, and highly reactive. The chief advantage offered over the multiple hearth approach is the pelletized product rather than the lime dust obtained with the multiple hearth.

8.4 Air Pollution Considerations

Incineration offers the opportunity to reduce sludge to a sterile landfill and remove offensive odors, but it also has the potential to be a significant contributor to the air pollution problem in an urban community. The quantity and size of particulate emission leaving the furnace of an incinerator varies widely, depending on such factors as the sludge being fired, operating procedures, and completeness of combustion. Incomplete combustion can form objectionable intermediate products, such as hydrocarbons and carbon monoxide.

National air pollution standards for discharges from municipal sludge incinerators have been promulgated which limit emissions of particulates (including visible emissions) from incinerators used to burn wastewater sludge as follows [40]:

1. No more than 0.65 g/kg dry sludge input (1.30 lb/ton dry sludge input).
2. Less than 20 percent opacity.

Visible emissions caused solely by the presence of uncombined water are not subject to the opacity standard.

Available data indicate that on the average, uncontrolled multiple hearth incinerator gases contain about 0.6 grain of particulate per standard cubic foot of dry gas [3]. Uncontrolled fluid bed reactor gases contain about 1.0 grain of particulate per standard cubic foot [13]. For average municipal wastewater sludge, this corresponds to about 33 pounds of particulates per ton of sludge burned in a multiple hearth, and about 45 pounds of particulates per ton of sludge burned in a fluid bed incinerator. Particulate collection efficiencies of 96 to 97 percent will be required to meet the standard, based on the above uncontrolled emission rate.

Sludge incinerators differ from most other types of incinerators in that the sludge does not supply enough heat to sustain combustion. Furthermore, there is less emphasis on retaining ash in the incinerator and much of it is discharged in stack gases. Particulate emissions to the atmosphere are almost entirely a function of the scrubber efficiency and are only minimally affected by incinerator conditions. Sludge incinerators in the United States are equipped with scrubbers of varying efficiency. These range from simple bubble-through type units to impingent type scrubbers with pressure drops up to 20 inches of water.

Existing state or local regulations in the United States tend to regulate sludge incinerator emissions through incinerator codes or process weight regulations [41]. Many state and local standards are corrected to a reference base of 12 percent carbon dioxide or 6 percent oxygen. Corrections to CO_2 or O_2 baselines are not directly related to the sludge incineration rate due to the high percentage of auxiliary fuel required. In some regulations, the CO_2 from fuel burning is subtracted from the total when determining compliance.

In developing the above standards, tests were conducted on the gaseous discharges from several sludge incinerators. Stack tests were conducted by EPA at five locations, including three multiple hearth incinerators and two fluid bed reactors as shown below [42]:

A. Fluidized bed reactor, 1,100 lb/hr dry solids design capacity, operated at 100 percent capacity during test, equipped with a 20 inch of water pressure drop venturi scrubber operated at 18 inches water pressure drop. Tested by EPA and by a state agency, latter using Code Method 8 (see Table 8-4 and Table 8-5).

B. Multiple hearth (six hearths) Herreshoff incinerator, 750 lb/hr dry solids design capacity, operated at 64 percent capacity during test, equipped with a 6.0 inch of water pressure drop single crossflow perforated-plate impinjet scrubber (see Table 8-6).

C. Multiple hearth (six hearths) Herreshoff incinerator, 900 lb/hr dry solids design capacity, operated at 35 percent capacity during test, equipped with a 6.0 inch water pressure drop single crossflow perforated-plate impinjet scrubber (see Table 8-7).

D. Fluidized bed reactor, 500 lb/hr dry solids design capacity, operated at 95 percent capacity during test, equipped with a 4.0 inch water pressure drop single crossflow perforated-plate impinjet scrubber (see Table 8-8).

E. Multiple hearth Herreshoff incinerator, 2,500 lb/hr dry solids design capacity, operated at about 50 percent capacity during tests, equipped with a 2.5 inch water pressure drop cyclonic inertial jet scrubber (see Table 8-9).

The results of these tests are shown in Tables 8-4 to 8-9. Figure 8-19 summarizes the results of the particulate measurements. The results from the unit using a venturi scrubber operating at 18-inch water pressure drop were used as the basis for the standard. The other systems using other types of scrubbers operating at lower pressure drops failed to meet the promulgated standard of 1.3 lb/ton dry sludge input. The study of these facilities indicated no relationship between the mass emissions and the percent of rated capacity at which the incinerator was operating, but a strong relationship between pressure drop across the scrubber and mass emissions was found [40]. All of the systems easily met the opacity standard. Observations at 15 other facilities indicated they all met a 10 percent opacity. The estimated costs of the scrubbing systems used as standard

TABLE 8-4

SLUDGE INCINERATOR FACILITY A_1 –SUMMARY OF RESULTS [42]

Run number	1	2	3	Average
Date	1-11-72	1-12-72	1-12-72	
Test time, minutes	108	108	108	108
Furnace feed rate, ton/hr dry solids	0.550	0.560	0.560	0.557
Stack effluent				
Flow rate, dscfm	2880	2550	2660	2700
Flow rate, dscf/ton feed	314,000	273,000	285,000	291,000
Temperature, ° F	59	59	59	59
Water vapor, vol. %	1.93	1.92	2.23	2.03
CO_2, vol. % dry	12.8	12.6	11.5	12.3
O_2, vol. % dry	4.8	4.7	6.4	5.3
CO, vol. % dry	0.0	0.0	0.0	0.0
SO_2 emissions, ppm	<0.3	<0.3	<0.3	<0.3
NO_x emissions, ppm	4.2	5.7	6.4	5.4
HCl emissions, ppm	<3.8	<2.9	<4.1	<3.6
Visible emissions, % opacity	<10	<10	<10	<10
Particulate emissions				
Probe and filter catch				
gr/dscf	0.024	0.005	0.004	0.011
gr/acf	0.023	0.005	0.004	0.011
lb/hr	0.583	0.116	0.099	0.266
lb/ton of feed	1.06	0.207	0.177	0.481
Total catch				
gr/dscf	0.032	0.007	0.010	0.0163
gr/acf	0.031	0.007	0.010	0.016
lb/hr	0.779	0.160	0.227	0.389
lb/ton of feed	1.42	0.286	0.405	0.704

TABLE 8-5

SLUDGE INCINERATOR FACILITY A_2 –SUMMARY OF RESULTS [42]

Run number	1	2	3	Average
Date	5-3-71	5-4-71	5-4-71	
Test time, minutes	60	60	60	60
Furnace feed rate, ton/hr dry solids	0.325	0.325	0.325	0.325
Stack effluent				
Flow rate, dscfm	3480	3600	3320	3470
Flow rate, dscf/ton feed	642,500	664,600	612,900	640,600
Temperature, ° F	80	80	78	79
Water vapor, vol. %	3.4	3.4	3.4	3.4
CO_2, vol. % dry (less aux. fuel)	4.0	5.1	4.0	4.4
SO_2 emissions[a]	–	–	–	–
Visible emissions Ringelmann No.[b]	<1	<1	<1	<1
Particulate emissions				
Total catch				
gr/dscf (cor. to 12% CO_2)	0.020	0.031	0.048	0.033
gr/acf	0.019	0.029	0.047	0.032
lb/hr	0.596	0.956	1.365	0.972
lb/ton of feed	1.84	2.94	4.20	2.99

Note: Tested by local agency using Code Method 1. Probe and filter catch not analyzed separately.

[a] No SO_2 detected.

[b] Opacity was not recorded.

TABLE 8-6

SLUDGE INCINERATOR FACILITY B–SUMMARY OF RESULTS [42]

Run number	1	2	3	Average
Date	10-13-71	10-14-71	10-14-71	
Test time, minutes	120	120	120	120
Furnace feed rate, tons/hr dry solids	0.237	0.236	0.249	0.241
Stack effluent				
Flow rate, dscfm	3300	2950	2120	2790
Flow rate, dscf/ton feed	835,000	750,000	511,000	699,000
Temperature, °F	198	196	199	198
Water vapor, vol. %	3.64	4.02	3.65	3.77
CO_2, vol. % dry	3.8	4.7	2.7	3.7
O_2, vol. % dry	17.3	1.40	15.8	15.7
CO, vol. % dry	0.0	0.0	0.0	0.0
SO_2 emissions, ppm	2.29 to 2.57	2.75	–	2.53
NO_X emissions, ppm	–	–	44.2 to 24.3 14.3	27.6
HCl emissions, ppm	–	–	0.624 to 1.33 0.621	0.858
Visible emissions, % opacity	<10	<10	<10	<10
Particulate emissions				
Probe and filter catch				
gr/dscf	0.0245	0.0196	0.0173	0.0205
gr/acf	0.0187	0.0155	0.0132	0.0158
lb/hr	0.690	0.495	0.315	0.500
lb/ton of feed	2.91	2.10	1.26	2.09
Total catch				
gr/dscf	0.0374	0.0374	0.0457	0.0402
gr/acf	0.0289	0.0287	0.0348	0.0308
lb/hr	1.06	0.945	0.832	0.946
lb/ton of feed	4.47	4.00	3.34	3.94

TABLE 8-7

SLUDGE INCINERATOR FACILITY C–SUMMARY OF RESULTS [42]

Run number	1	2	3	Average
Date	7-15-71	7-15-71	7-16-71	
Test time, minutes	80	80	80	80
Furnace feed rate, tons/hr dry solids	0.111	0.149	0.146	0.135
Stack effluent				
Flow rate, dscfm	1230	1490	1400	1373
Flow rate, dscf/ton feed	665,000	600,000	575,000	613,000
Temperature, °F	80	80	77	79
Water vapor, vol. %	3.23	3.00	2.95	3.06
CO_2, vol. % dry	10.0	10.1	10.2	10.1
O_2, vol. % dry	7.7	7.3	7.4	7.5
CO, vol. % dry	0.0	0.0	0.0	0.0
SO_2 emissions, ppm	15.9 to 11.9	14.5 to 14.6	14.6 to 13.3	14.2
NO_X emissions, ppm	402 to 140	90.8 to 74.3	14.5 to 142 50.6 to 61.8	163
HCl emissions, ppm	3.50 to 2.62	2.33 to 2.62	2.52 to 2.62	2.72
Visible emissions, % opacity	<10	<10	<10	<10
Particulate emissions				
Probe and filter catch				
gr/dscf	0.0127	0.0620	0.0196	0.0314
gr/acf	0.00985	0.0477	0.0152	0.0242
lb/hr	0.127	0.620	0.196	0.314
lb/ton of feed	1.14	4.16	1.34	2.21
Total catch				
gr/dscf	0.0195	0.0696	0.0260	0.0384
gr/acf	0.0150	0.0535	0.0201	0.0295
lb/hr	0.206	0.889	0.312	0.469
lb/ton of feed	1.86	5.97	2.14	3.23

TABLE 8-8

SLUDGE INCINERATOR FACILITY D–SUMMARY OF RESULTS [42]

Run number	1	2	3	Average
Date	7-21-71	7-21-71	7-22-71	
Test time, minutes	120	96	96	104
Furnace feed rate, tons/hr dry solids	0.255	0.237	0.202	0.231
Stack effluent				
Flow rate, dscfm	1190	1170	1240	1200
Flow rate, dscf/ton feed	280,000	296,000	368,000	315,000
Temperature, °F	99	99	95	98
Water vapor, vol. %	3.92	4.90	3.48	3.83
CO_2, vol. % dry	8.8	9.9	9.1	9.3
O_2, vol. % dry	6.3	7.4	8.2	7.3
CO, vol. % dry	0.0	0.0	0.0	0.0
SO_2 emissions, ppm	8.29 to 11.2	14.8 to 14.8	14.2 to 15.4 17.8	13.8
NO_X emissions, ppm	154 to 168	41.2 to 42.9	187 to 170 161	132
HCl emissions, ppm	0.780 to 260	4.16 to 1.56	2.35 to 2.09	2.26
Visible emission, % opacity	<10	<10	<10	<10
Particulate emissions				
Probe and filter catch				
gr/dscf	0.0551	0.0766	0.0545	0.0621
gr/acf	0.0468	0.0650	0.0467	0.0528
lb/hr	0.562	0.768	0.579	0.636
lb/ton of feed	2.20	3.24	2.87	2.77
Total catch				
gr/dscf	0.0665	0.0859	0.0653	0.0726
gr/acf	0.0565	0.0729	0.0559	0.0618
lb/hr	0.678	0.861	0.694	0.744
lb/ton of feed	2.66	3.63	3.43	3.24

TABLE 8-9

SLUDGE INCINERATOR FACILITY E–SUMMARY OF RESULTS [42]

Run number	1	2	3	Average
Date	8-5-71	8-5-71	8-5-71	
Test time, minutes	96	96	96	96
Furnace feed rate, tons/hr dry solids	0.689	0.855	0.290	0.611
Stack effluent				
Flow rate, dscfm	9840	8510	10,290	9547
Flow rate, dscf/ton feed	–	–	–	–
Temperature, °F	135	145	145	142
Water vapor, vol. %	16.3	18.6	14.8	16.6
CO_2, vol. % dry	4.2	4.3	2.2	3.6
O_2, vol. % dry	14.9	14.9	16.9	15.6
CO, vol. % dry	0.0	0.0	0.0	0.0
SO_2 emissions, ppm	2.01	2.07	2.12	2.07
NO_X emissions, ppm	62.8 to 46.0	83.5 to 75.8	44.3 to 54.7	61.2
HCl emissions, ppm	11.9	6.83	10.9	9.88
Visible emissions, % opacity	<10	<10	<10	<10
Particulate emissions				
Probe and filter catch				
gr/dscf	0.0260	0.0136	0.0134	0.0177
gr/acf	0.0196	0.0099	0.0101	0.0132
lb/hr	2.19	0.99	1.18	1.45
lb/ton of feed	3.18	1.16	4.07	2.80
Total catch				
gr/dscf	0.0335	0.0221	0.0170	0.0242
gr/acf	0.0252	0.0159	0.0128	0.180
lb/hr	2.83	1.61	1.50	1.98
lb/ton of feed	4.11	1.88	5.17	3.72

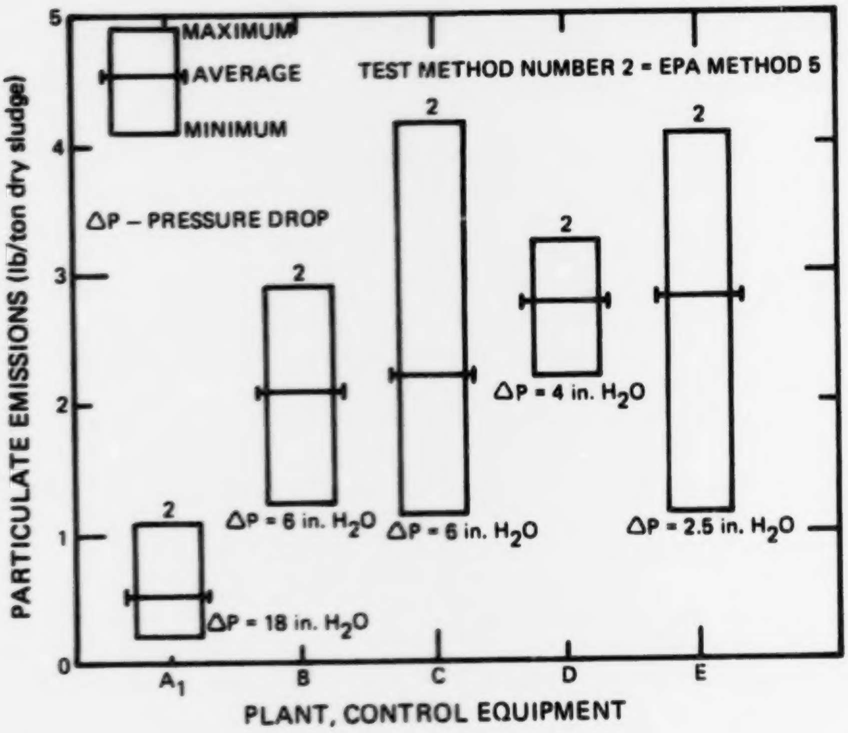


FIGURE 8-19. Particulate emissions from sludge incinerators at wastewater treatment plants [40].

practice in the U.S. are typically about 4 percent of the total incineration facility for a plant serving 100,000 people. The scrubber required to achieve the proposed particulate standards would increase the cost an estimated 0.4 percent. Annual operating costs were estimated to be increased by 0.9 percent [41].

Wastewater sludges contain metals which could be toxic if discharged into the atmosphere. Unfortunately, there are very few data on the metals being discharged to the atmosphere from municipal sludge incineration. The forms in which metals are found in sludge will influence their behavior on incineration [41]. For example, if cadmium is present in the sludge in solution as cadmium chloride, it could volatilize upon incineration. If it is present as a precipitated hydroxide, it would probably decompose to the oxide, but would not volatilize at the temperatures of incineration. However, it is believed that most of the toxic metals, with the exception of mercury, will not disproportionately appear in stack gases because of volatilization, but will be converted to oxides and appear in the particulates removed by scrubbers or electrostatic precipitators and in the ash.

Gaseous pollutants which could be released by sludge incineration are hydrogen chloride, sulfur dioxide, oxides of nitrogen, and carbon monoxide. Data are presented in Tables 8-4 to 8-9 on the quantities of these materials found in stack gases. Carbon monoxide is no threat if the incinerator is properly designed and operated. Hydrogen chloride, which would be generated by decomposition of certain plastics, is not a significant problem at concentrations currently observed. Consideration of the possibility of SO_2 and NO_X pollution is aided by examination of the sulfur and nitrogen content of sludges. Sulfur content is relatively low in most sludges. In addition, much of this sulfur is in the form of sulfate, which originated in the wastewater. Sulfur dioxide is not expected to be a serious problem. Sludge typically has a high nitrogen content from proteinaceous compounds and ammonium ion. Limited data are available for predicting whether a high proportion of these materials will be converted to oxides of nitrogen on combustion. From the data available, the concentration of oxides of nitrogen from sludge incineration should be less than 100 ppm from a properly operated incinerator and were observed to be less than 10 ppm from Facility A (Table 8-4). Considering this low concentration, the production of oxides of nitrogen will probably not limit the use of incineration for disposing of sludge in most cases.

Some data are available on the discharge of manganese and nickel which indicate the following emission rates from incinerator facilities with emission controls:

Metal	Emission Factor Range Millipounds/ton (dry solids)
Mn	0.5 to 1.6
Ni	0.2 to 8.2

In both cases, the higher limit is represented by a single measurement.

Mercury is an example of a substance which presents special problems during incineration. High temperatures during incineration decompose mercury compounds to volatile mercuric oxide or metallic mercury. Fortunately, the quantity of mercury involved is small. Based on an estimate that approximately 4,000 tons per day of sludge are incinerated, and an average Hg concentration in sludge of 0.01 mg/g, 80 pounds of mercury would be expelled into the atmosphere over the United States. This amount compares to the estimated 3,000 tons per day of mercury which is discharged into the atmosphere from the burning of coal over the earth [41]. Limited test data [43] indicate that perhaps only 4 to 35 percent of the mercury entering an incinerator with emission controls will be emitted to the atmosphere (excluding particulate forms).

In addition to the major air pollutants resulting from the burning of sludge, toxic substances can arise due to the content of pesticides or other organic compounds in the sludge. Unfortunately, very limited data are available on the concentrations of these materials in municipal sludges or their fate in an incinerator. Data reported by EPA [41], in a random selection of sludges, showed the following levels of materials present in the raw sludges:

Compound	Range (ppm)
Aldrin	16 (in one sludge only)
Dieldrin	0.08 to 2.0
Chlordane	3.0 to 32
DDD	not detected to 0.5
DDT	not detected to 1.1
PCB's	not detected to 105

Pesticide and PCB determinations were made on sludges collected during the incinerator tests at three of the five plants listed in Tables 8-4 to 8-9 [41]. PCB's were found in all of these sludges, but concentrations were low (1.2 to 2.5 ppm). Pesticides and PCB's were found only in the sludge. They were not found in the ash from either type incinerator, nor in the inlet or outlet scrubber water. Ash can be analyzed for these materials to the same degree of sensitivity as the sludge. A level of 0.1 microgram/g (ppm) could have easily been detected. It is quite certain that these materials are not being carried out in the ash.

The mass flow rate of water to the scrubber is about 400 times the dry solids flow rate to the incinerator. Consequently, the concentration at which these materials can be detected in water must be sufficiently low to be sure that they are not escaping in the scrubber water. Fortunately, analytical techniques are such that these materials can be detected in water down to 0.1 nanogram/g (ppb). Thus, it is reasonable to believe that they are not in the scrubber water.

Since the PCB's do not appear in ash or scrubber water, they are either destroyed by incineration or remain as vapors in the water-scrubbed (and cooled) gas stream. All of these materials have some solubility in water and it is likely that no trace would be present in the scrubber water. Consequently, their escape as vapors from the incinerators seems unlikely. However, one should also examine the available data on the decomposition of PCB's and pesticides in other situations.

Rapid thermal degradation of most pesticides has been shown to begin at approximately 500° C with near total destruction at 900° C (1,652° F) [44,45]. If these materials volatilize before burning, the use of afterburners on incinerators would be needed to provide complete destruction. One manufacturer of pesticides achieves near total destruction of pesticides in its multiple hearth carbon regenerator by providing an afterburner using a 0.23 to 0.80 second retention time at 1,600 to 1,800° F.

The PCB's are even more thermally stable than most pesticides, as one would suspect. An incinerator at St. Louis, Missouri, achieves total destruction of concentrated PCB's at 2,400° F with a retention time of 2.5 seconds. Experiments have shown, however, that 99 percent destruction is possible at 1,600 to 1,800° F in 2.0 seconds.

A privately funded study [46] found that PCB was completely destroyed in a multiple hearth furnace burning organic sludges when the exhaust gas temperature was 1,100° F and that 94 percent destruction occurred at normal exhaust temperatures.

The EPA Sewage Sludge Incineration Task Force [41] concluded that it has been adequately demonstrated that existing well-designed and operated municipal wastewater sludge incinerators are capable of meeting the most stringent particulate emission control regulation existing in any state or local control agency. This observation coupled with the fact that the newly promulgated federal standards are based on demonstrated performance of an operating facility indicates that use of proper emission controls and proper operation of the incineration system will enable a facility to meet all existing air pollution regulations. Although only the venturi scrubber (Tables 8-4 to 8-9) met the promulgated standard in the EPA tests. EPA [40] has stated that:

> Impingement scrubbers tested by EPA did not meet the standard but, in our best judgment, would do so if used in conjunction with an oxygen meter that automatically regulates fuel burning rate. In our best judgment, electrostatic precipitators could also provide more than adequate control. There are no EPA test data on either of these control systems because during the test program there were no existing plants using them.

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CHAPTER 9

FINAL DISPOSAL PROCESSES

9.1 Methods, Functions, and Occurrences

The need for proper treatment of all residual sludge streams is well recognized and should be provided for in plant design. Lakes and streams are no longer acceptable for sludge disposal. The methods of final disposal can be broadly categorized as disposal or utilization procedures. Disposal procedures include landfills and ocean dumping as described in Table 9-1.

TABLE 9-1

FINAL DISPOSAL METHODS

Disposal Procedure	Principal Sludge Form	Main Constraints
Sanitary Landfill	Dewatered cake or ash (Stabilized)	Gas leachate, and runoff control; land availability
Ocean Dumping	Liquid (Thickened)	Oceanic and shoreline pollution

The landfill is one of the major methods for final disposal of sludge and incinerator ash. The amount of sludge that is dumped in the ocean is decreasing because of regulations.

Utilization procedures are receiving increasing attention and are described in Table 9-2. Cropland application and land reclamation are the major sludge utilization methods. The EPA municipal inventory shows that landfill and land spreading of sludge are used in 50 percent of the U.S. installations. Proceedings of recent symposia [1,2] on this topic are available and contain much useful information. Other methods of synthesizing or retrieving useful products by treating sludge with chemicals and/or heat are being studied.

TABLE 9-2

SLUDGE UTILIZATION METHODS

Utilization Procedure	Sludge Form	Main Constraints
Cropland Application	Liquid, cake dried, or compost	Application rate, unsatisfactory sludge
Land Reclamation	Liquid or Dewatered	Application rate, unsatisfactory sludge, availability of land

9.2 Selection of Method of Final Disposal

In developing an optimum conceptual design for a wastewater treatment plant, it is increasingly apparent that determination of the method of final sludge disposal is a major consideration. The method of final disposal determines the acceptable form of the sludge residue and thus influences the choice of both sludge and liquid unit processes to be employed.

In selecting a final disposal or utilization process, the method chosen should be in accordance with local, state, interstate, and federal requirements. While no sludge residues, grit, ash, or other solids should be discharged into the plant effluent or receiving waters, the procedure chosen should also not result in any significant degradation of surface or groundwater; air or land surfaces.

Since present indications are that ocean disposal could be banned, designs incorporating sludge disposal to the ocean are questionable. It is desirable that methods chosen do not cause a health hazard or a nuisance condition, therefore it is essential that sludges be stabilized prior to spreading on land. The acceptable form of sludge for disposal or utilization is partially determined by the sludge treatment method. The sludge may be in the form of a liquid, dewatered cake, incinerator ash, compost product, or dried powder.

9.3 Sanitary Landfill

Stabilized sludge containing no free water can be satisfactorily disposed in a sanitary landfill either alone or in a mixture with municipal solid waste. A sanitary landfill must be managed so that wastes are systematically deposited and covered with earth to control environmetal impacts within defined limits. This distinguishes a sanitary landfill from an uncontrolled dumping operation. Sludges and solid wastes have in the past been disposed of in dumps not meeting proper landfill specifications. The placement of incinerator ash or stablized sludge cake in a sanitary landfill can be an acceptable procedure when adequate land is available and site location and operational precautions prevent the creation of nuisance conditions or health hazards. Prior to placing sludge in a landfill it should be sufficiently dewatered to minimize the quantity of free water present. Leachate and runoff from a sanitary landfill should be minimized and when necessary collected and suitably treated to prevent pollution of ground and surface waters. Therefore, sound engineering judgment dictates that sanitary landfills not be located in an existing flood plain.

9.3.1 Design Criteria

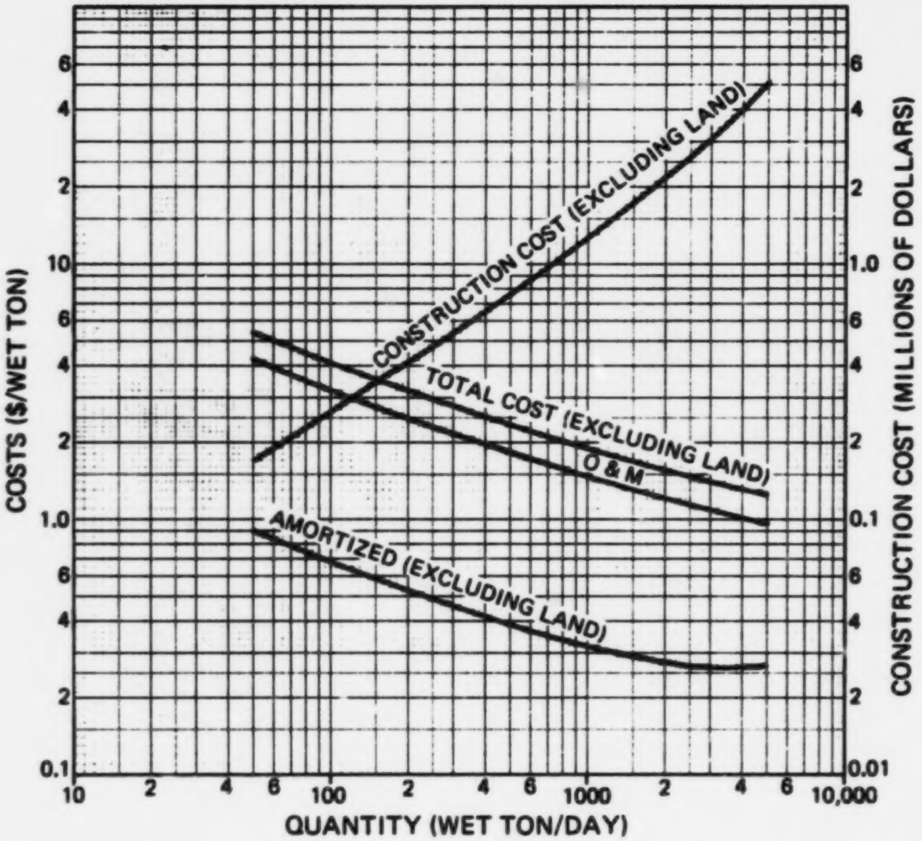
Wilcomb and Hichman [3] suggest that the site of a sanitary landfill be easily accessible and safeguarded against uncontrolled gas movement from the decomposition of organic matter. During the predesign survey, the site's geology, hydrology, and soil conditions should be considered relative to the need for adequate protection of groundwater, conformation of area land use planning, and provision of an adequate quantity of earth cover.

The landfill itself should have limited access and provision for uniform spreading of wastes in layers not over 2 feet thick, followed by compaction. The compacted wastes should have a minimum of six inches of suitable compacted earth cover at the end of each working day. When each portion of the landfill is completed, a uniform layer of earth cover compacted to a minimum depth of 2 feet should be placed over it and suitable grasses planted to prevent erosion.

Adequate monitoring of any land application or landfill site is essential. This plan must be specifically designed for applicable local conditions and should include monitoring groundwater observation wells, surface water, sludge and soils for heavy metals, persistent organics, pathogens, and nitrates. Human food chain products grown in sludge aided soils should also be monitored for heavy metals, persistent organics, and pathogens.

9.3.2 Costs of Sanitary Landfill

Even with increases in costs resulting from more stringent environmental impact control procedures, landfilling as practiced is generally less expensive than other final disposal or utilization procedures. An important consideration can be the dewatering requirement. Further, land costs can be a major factor. Figure 9-1 [4] presents a range of capital and operation and maintenance costs for landfills, excluding land costs. Quantities are expressed in tons of wet sludge cake per day. Other investigators have reported landfilling costs of from $1 to $4 per ton of dry solids [5].



NOTES:

1. Minneapolis. Mar., 1972. ENR Construction Cost Index of 1827.
2. Amortization of 7% for 20 years.
3. Labor rate of $6.25 per hour.
4. Quantity assumes 6-day work week.
5. Wet sludge must be considered for cost per ton.
6. Source: U. S. P. H. S. and Stanley Consultants.

FIGURE 9-1. Capital and O/M costs for sanitary landfills [4].

9.4 Use of Sludge on Agricultural Land

The application of sludge to farmland is a very popular utilization method because it can be both economical and simple. Limitations include heavy metals, occasional public resistance, and the unavailability of suitable land. The popularity of sludge spreading for crop production and soil improvement has increased significantly over the last 10 years.

9.4.1 Soil Considerations

Soil is composed of mineral matter, organic matter, microorganisms, solutions, and air. The soil's assimilative capacity hinges on its ability to filter, buffer, and absorb a sludge's constituents. It chemically and biologically transforms materials and supports plants which use the applied nutrients. Desirable soil properties for sludge assimilation are:

- Depth
- High infiltration and percolation capacity.
- Fine enough texture to have high water and nutrient holding capacity.
- Good drainability and aeration.
- Neutral or alkaline pH.

9.4.2 Sludge as a Fertilizer and Soil Conditioner

Municipal sludge contains all of the essential plant nutrients. Almost half of the nitrogen and potassium in digested sludge is in the liquid phase, so drying or dewatering can decrease these nutrients significantly. The ratio of potassium to nitrogen and phosphorus in sludge is low in relation to crop needs. Therefore, the use of digested sludge at a rate required to just supply the nitrogen needs of a crop will usually not supply enough potassium. Table 9-3 presents the nutrient content for a liquid digested mixture of primary and activated sludges.

TABLE 9-3

PRIMARY NUTRIENT CONTENT OF LIQUID DIGESTED SLUDGE
(25 Percent of Dry Weight)

Sludge Type	Nitrogen (%)	Phosphate (P_2O_5) (%)	Potash (K_2O) (%)
Liquid Digested	3.5 - 6.4	1.8 - 8.7	0.24 - 0.84

Dotson [6] indicated a 2-inch application of Chicago's liquid digested sludge to cropland would supply about 200 to 350 lb of ammonia nitrogen, about the same amount of organic nitrogen, 250 to 400 lb of phosphorus (approximately 80 percent organic), and about 60 lb of potassium per acre. A good corn crop can utilize 150 to 250 lb or more of nitrogen per acre, and some grasses can use more, but increased leakage of nitrogen may occur at these higher loadings. Crop removal and volatilization determine how much of the soluble soil nitrogen will be leached from the soil. The principal sludge components that determine satisfactory application rates of sludge to cropland are nitrogen, trace elements, and pathogens.

Nitrogen is typically the first sludge component to limit the rate of sludge application to land, since adding excess nitrogen to soil involves the very real risk of polluting the groundwater with nitrates.

High nitrate concentrations are toxic to human and livestock. Public Health Service and the World Health Organization drinking water standard for nitrate nitrogen is 10 mg/l. If nitrate nitrogen is applied in amounts greater than can be removed by harvested plant uptake, the excess nitrates can potentially contaminate ground and/or surface waters by leaching or runoff, respectively. Furthermore, excess nitrogen can also cause a high nitrate fodder yield in certain crops. Methemoglobinemia in infants can occur from consuming green forage silage which contains in excess of 3.0 percent nitrate (dry basis).

The pathways of nitrogen in soils and the hazards they present to groundwater are difficult to predict. The processes that affect the form of nitrogen in soils (mineralization, nitrification, denitrification, immobilization, fixation, adsorption, volatilization, cation exchange, convection, dispersion, and plant uptake) may take place concurrently, and the rates at which they progress are determined largely by soil type and climate.

Mineralization (the conversion of organic nitrogen to ammonia) proceeds at variable rates depending on climate and soil conditions and the nature of the organic matter, and nitrification (the oxidation of ammonia nitrogen to nitrate nitrogen) is relatively fast in aerobic soils with favorable temperature. On the other hand, denitrification (transformation of nitrate nitrogen to nitrogen gas) takes place where free oxygen is absent or deficient and where other conditions, including a supply of carbon, are favorable for biological activity.

Microbes utilize part of the available nitrogen in soils to synthesize new cells. Ammonia ions may be fixed by organic matter and silicate clays and protected from biological attack (fixation). Volatilization of ammonia may be substantial from soils with high pH. Plant uptake varies greatly, of course, and the amount of nitrogen removed in runoff varies with precipitation patterns and farming practices.

A hypothetical model indicates that only a little more than one-half inch of digested sludge per year could be applied without contributing nitrogen to leachates [7].

However, at the Agricultural Research Center in Beltsville, Maryland, adding up to 160 tons of sludge solids per acre to soils with clay subsoils did not cause nitrates to leach into groundwater at a depth of 9 feet. Ammonia compounds in excess are toxic to some seeds. Two weeks cessation of sludge application before seeding has been adequate to avoid this problem.

It is difficult to formulate guidelines as to the rate that sludge can be applied to cropland without nitrate pollution of groundwater. Soil type, geology, climate, crops, and farm management are important factors in determining the fate of nitrogen added in sludge. Careful calculations should be made for any proposed system based upon applicable local conditions.

Metal content of sludge varies widely. The metals of most concern are zinc, copper, nickel, cadmium, and lead. Zinc and copper are micronutrients that may enhance crop quality when sludge is spread thinly. The heavy metal content of sludge can vary widely from treatment plant to treatment plant as illustrated in Table 9-4.

TABLE 9-4

HEAVY METAL CONTENTS IN SLUDGE [6]
(mg/l, Dry Basis)

Location	Zinc	Copper	Nickel	Cadmium
Dayton, Ohio	8,390	6,020	<200	830
Monterey, California	3,400	720	220	<220
Tahoe, California	1,700	1,150	<400	40
Millcreek, Cincinnati, Ohio	9,000	4,200	600	<40

It has been said that the heavy metal content is a function of industrial wastes and that sludges from treatment plants not serving large amounts of industry should not have this problem. Unfortunately, analysis of sludges from such cities as Washington, D.C., does not support this supposition. Some sludges of apparently predominantly domestic origin contain significant heavy metal concentrations. Prior to land spreading of sludges, analysis of the influent to the treatment plant should be made. If high heavy metal concentrations are found, sound engineering procedures to remove these metals at the source should be initiated.

Allowable levels of trace elements in soils are difficult to establish because of the many complex soil and crop factors. Plants vary widely by species, varieties, and strains in tolerance to trace elements. Interdependent soil properties greatly influence the availability of trace elements to plants. Thus, spreading wastewater sludge on cropland, like other disposal methods, involves an element of risk.

Usually trace element toxicities to plants are more prevalent and acute where sludges are applied to acid soils. Also plant species exhibit rather marked differences in tolerance to levels of trace elements in sludge-amended soils. The amount of sludge which can be safely applied to soil, will depend upon the composition of the sludge, the kind of soil to which it is applied, and the species of plant grown on the soil. Detailed discussions of specific trace elements and their effects are available in the literature [8,1,2].

Pathogen control is of importance because of the possible exposure directly to sludge in the handling and application steps and in the food chain. Although anaerobic digestion reduces the pathogen content of sludge, a significant number of pathogens may survive the process. Pathogens in sludge can be destroyed through:

- Storing for long periods.
- Pasteurizing at 70° C for 30 minutes.
- Adding lime to raise pH to 12.4.
- Using chlorine to stabilize and disinfect sludge.
- Using other chemicals.

Stabilization techniques are covered in Chapter 5. Long storage of wastewater sludge has been suggested as one of the simplest methods of reducing pathogenic organisms. Storing sludge for 30 days has reduced fecal coliforms by 99.9 percent, although some parasites probably persist much longer when sludge is stored in lagoons. Most large municipalities that dispose of sludge by spreading on land store it in lagoons–this provides the flexibility needed at times when sludge cannot be spread. Access to storage lagoons should be restricted. Sludge spread on pastures during the grazing season in Germany and Switzerland is pasteurized. Maintaining a temperature of 70° C for 25 to 30 minutes kills pathogens, viruses, cysts, worm eggs, and oocytes. Direct steam injection avoids fouling and sealing of heat exchangers. Heat recovery is uneconomical for small plants, but larger plants may use heat exchange to minimize energy requirements.

Aerosols which could contain pathogenic microorganisms may be present in the air over a landspreading site. Where spray irrigation is used to distribute the sludge, the potential for aerosols is increased.

Three methods have been proposed to be used (either concurrently or separately) to minimize aerosols: (1) low pressure spray, (2) downward directed spray, and (3) large droplet spray. The potential for erosion should be checked before using these methods. Application methods other than spraying minimize aerosols. At many reclamation sites, however, spraying is the only practical method. Where spraying is used, spraying methods should be selected which will minimize aerosol formation and wide buffer zones should be used. Mosquito vectors can be controlled by eliminating ponding on the site. This starts with properly grading the site and is made effective by maintenance to correct minor ponding locations and prevent others from forming.

Fruits and vegetables grown on wastewater irrigated or sludge-amended soils can be surface-contaminated with pathogenic microorganisms. The viability of these pathogens is extremely variable and may be from a few hours to several months. Also, among the factors which influence the survival of pathogens in the soil and on vegetation are:

- Type of organism.
- Temperature - lower temperature increases viability.
- Moisture - longevity is greater in moist soils than in dry soils.
- Type of soil - neutral, high moisture holding soils favor survival.
- Organic matter - the type and amount of organic matter present may serve as a food or energy source to sustain the microorganisms.

The fact that some pathogens can survive the sludge digestion process, and remain viable in the soil for periods up to several months has prompted the general recommendation that liquid sludges should not be applied to root crops or crops intended for human consumption in the raw form.

Pastureland and farmland used to grow forage crops are frequently used as land disposal sites, a practice which appears to present little problem from the standpoint of disease transmission via livestock grazing on such fields.

The potential for contamination of groundwater sources by pathogenic microorganisms is dependent on the ability of the pathogens to survive and move through the soil system. A number of factors including the types of soil involved, the number of microorganisms applied to the soil surface, and the many different kinds of microorganisms contained in both the soil system and the applied wastes all combine to complicate an assessment of pathogen movement in soils. Fine clay soils are more effective than sandy soils for removal of pathogens. A soil system is generalyy efficient in removing pathogens unless movement of sludge or effluent through faults occurs. It has been reported that bacteria do not travel horizontally more than 100 feet through granular soils, that complete removal of *E. coli* was noted in about 16 feet of dune sands, and viruses were removed in a 2-foot bed of clean sand at moderate rates of application over a 7-month period. Generally, groundwater contamination by pathogens may not be as serious a potential hazard as surface water contamination via prolonged surface erosion and/or either direct runoff from the sludge spreading operation or surface water runoff induced by snowmelt or rainfall. With careful site management practices, the problem of surface runoff can be minimized. Certainly, each site must be carefully evaluated.

9.4.3 Physical Process Considerations

In developing a system for the application of sludge to cropland, the mode of transportation, application procedure, and rate of application must be considered. Transportation may be accomplished by tank truck, barge, rail, or pipeline. Sludge characteristics, elevation differences, distance, sludge volume, and land availability are important factors in selecting a method of transporting sludge from the treatment plant to the utilization site. Tank trucks afford flexibility in the selection of utilization sites, and they are widely used to haul and apply sludge. The ton to mile cost is relatively high, so small communities with available land near the treatment plant are most apt to find the use of the tank trucks feasible. Pipelines usually entail relatively high capital and low operating costs, so assurance of the availability of land for a long period of time is an important consideration. Cost analysis should be used in selecting the mode of transportation. Figure 9-2 [7] shows relative transportation costs for liquid organic sludges.

Retention time at the treatment plant or storage facilities near the land application site can provide for periods, when sludge spreading is not feasible. Storage also diminishes the pathogen population and further stabilizes the sludge.

Ridge and furrow irrigation methods are used for applying sludge to the land surface. These methods are better adapted to level land and cold climates. They also minimize potential air-borne virus and aesthetic problems. Spray sprinkler irrigation systems however, are more flexible, require less soil preparation, and can be used with a wider variety of crops than the ridge and furrow methods. Figure 9-3 shows such a system. In addition to surface sludge application several systems have been developed for incorporating sludge of from 1 to 85 percent solids into the soil. Incorporating sludge in the soil is a safeguard against odors, aesthetic problems, contamination of surface waters from erosion, or other problems that can result from irrigation techniques.

Application rates depend on sludge composition, soil characteristics, climate, vegetation, and cropping practices. Annual application rates have varied from 0.5 to more than 100 tons per acre. Applying sludge at a rate to support the nitrogen needs of a crop, usually about 5 to 10 tons of digested sludge solids in the liquid form, avoids problems associated with overloading the soil. A rough guide for selecting the application rate for an acceptable sludge is given in Table 9-5.

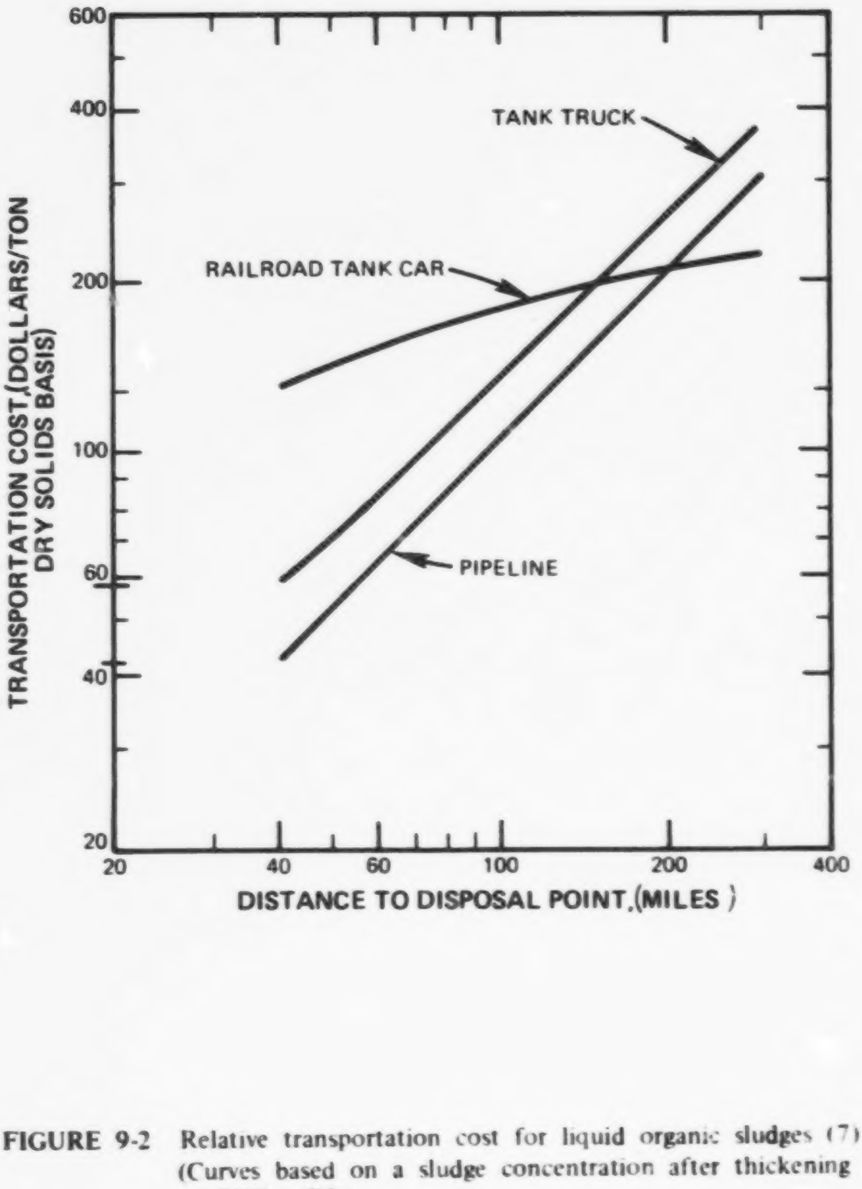


FIGURE 9-2 Relative transportation cost for liquid organic sludges (7) (Curves based on a sludge concentration after thickening to 3½% solids)

FIGURE 9-3. Typical spray sprinkler.

TABLE 9-5

APPLICATION RATES TO CROPLAND

Relative Soil Conditions	Application Rate (tons dry sl. solids/acre/year)
Slight Limitations	10 - 20
Moderate Limitations	<10

Specific points which must be continuously considered and/or monitored during utilization are:

- The trace element composition of sludge, soil, and crops.
- The nitrogen content of sludge, soil, and crops, and potential nitrate pollution of aquifers.
- The use of only disinfected sludge on low-growing fruits or vegetables to be consumed raw.
- The hydraulic overloading of soil.
- The ultimate use of land.
- The practices to control runoff and erosion.

9.4.4 Crop Considerations

Crops vary widely in their reaction to sludge enriched soils. The particular crop species may be adversely affected by trace elements in the sludge. Additionally, the crop may take up and concentrate certain of these trace elements, thereby, inhibiting future use of the harvested materials (particularly in the human food chain). The reaction of a specific crop to sludge application is extremely site dependent. Factors such as soil type, pH, moisture content, climate, and crop species are important. For advice concerning crops which can be satisfactorily grown in sludge enriched soils, the local representatives of the U.S. Department of Agriculture should be consulted.

To date, the practice of sludge spreading in forests has been limited. However, forests offer opportunities for beneficial use of sludge to improve soil fertility and increase tree growth. With most tree species nutrient uptake is small compared to that of cultivated crops, however in the case of some species, intensive culture operations for the production of wood fibre is possible. High application rates might require nitrogen removal to prevent nitrate pollution of groundwater.

9.4.5 Costs of Cropland Sludge Spreading

Table 9-6 presents actual costs data for land spreading of digested sludges. As can be noted, the economics of this final utilization procedure can vary widely.

TABLE 9-6

COSTS FOR LAND SPREADING DIGESTED SLUDGE

Location	Approximate Plant Size (mgd)	Cost ($/Ton)[a]	Reference
Chicago, Illinois	1,300	60.54 = Current 35.24 = Ultimate[b]	[9]
San Diego, California	90	10.57	[10]
Piqua, Ohio	3.8	17.50 - 30.00	[10]
St. Marys, Pennsylvania	1.3	19.92	[11]

[a] Excludes digestion and costs are given per ton of digested solids at 1972 prices.

[b] Ultimate costs include pipeline to be constructed. A principle variable is transportation cost.

9.5 Land Reclamation

A large part of the previous material on cropland utilization also applies to deposition on land for reclamation purposes. High application rates are commonly used to reclaim strip-mine spoils or other low-quality land and may lead to water contamination if drainage and runoff controls are not installed. Leachates that are unsuitable for groundwater recharge may be intercepted by tile drains and treated before being released to the environment. Surface runoff can also be impounded and treated. The great amount of sludge sometimes used to reclaim land may cause accumulations of trace elements in excess of normal concentrations in soils, however, soil management can maintain high organic matter content. A nearly neutral pH minimizes toxicity to plants.

9.6 Land Disposal Case Studies

As the pace of activity in this class of methods increases, more actual use data become available. The periodical, *Compost Science* is a particular focal point for such information.

9.6.1 St. Marys, Pennsylvania

The St. Marys, Pennsylvania treatment plant is a 1.3 mgd activated sludge facility treating principally domestic wastewater. Anaerobically digested sludge with a 5 percent solids concentration was applied at the rate of 900,000 gallons/year to primarily cropland for the past 13 years. Loading of a 1,500-gallon capacity tank truck involved a 10-minute fill-up period and about 5 minutes for spray application. The average round trip was 5 to 8 miles. Sludge was spread on hay, pasture, oats stubble, corn stubble, poor lawn, brush, orchards, and athletic fields. Figure 9-4 and 9-5 show a typical tank truck operation and equipment. The average application rates for two of the fields is shown in Table 9-7.

TABLE 9-7

APPLICATION RATES AT ST. MARYS

Field	% Solids	gal/acre/yr	Tons Dry Solids acre/year
Hay	4.25	19,600	3.54
Pasture	3.7	17,450	2.60

No complaints arose in St. Marys, and farmers welcomed the sludge. Cost of disposal averaged $19.92/ton [7].

9.6.2 Fergus Falls, Minnesota

The Fergus Falls wastewater treatment plant processes 1.6 mgd (76 percent) domestic wastewater and 0.5 mgd (24 percent) industrial wastewater. Sludge is produced by sedimentation and high rate trickling filters and is anaerobically digested. At one time, sludges were dewatered on drying beds, but these were removed in 1959. Since 1959, the digested sludge has been spread on farm fields in the summer and a city owned golf course during the winter. The secondary digester acts as a holding tank for the sludge. Sludge is hauled and applied with a 2,200-gallon capacity tank truck, and normally six

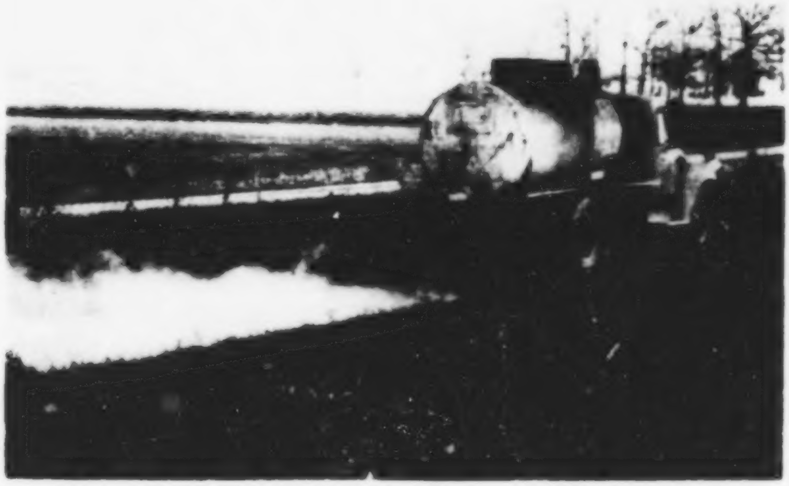


FIGURE 9-4. Tank trunk spreading sludge in cold weather.

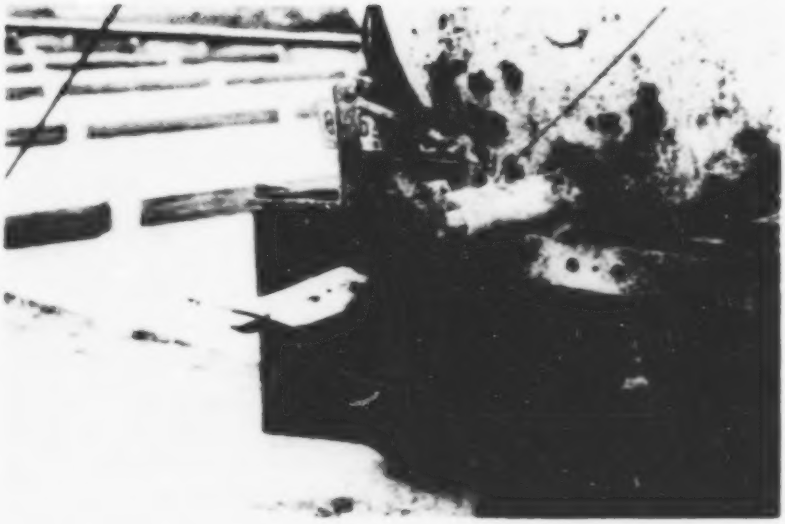


FIGURE 9-5. Close-up view of sludge deflection plate.

loads per day are hauled by one full-time operator. The solids content of the sludge is normally about 3.5 percent. In the summer, the sludge is applied to about 45 acres; twenty acres of pasture and 25 acres that are planted in corn the following year. The application is made on alternative years on the pasture land and the cornfield. The hauling distance is about 7 miles to the golf course and about 6 miles to the farm [8].

9.6.3 Xenia, Ohio

Landspreading of liquid sludges has been utilized by Xenia's two activated sludge wastewater treatment plants for about 13 years. Both anaerobically and aerobically digested sludges are being disposed of by land spreading [8]. Typically, the sludges average about 2.3 to 3.5 percent solids at 69 to 74 percent volatile solids and have an average pH of 7 to 7.2. The influent to these plants is essentially all domestic wastewater. Although a vacuum filter has been installed at the Glady Run plant and provisions have been made for a vacuum filter at the new Ford Road plant, land spreading continues to be the method of sludge disposal.

Presently, the sludge is spread on farmland used to grow field corn, wheat, and forage crops, and on pastureland used for cattle grazing. During the growing season, the liquid sludge is hauled extensively to pasturelands with the cattle remaining in the fields during the spreading operation. At the present time, 10 farms are being utilized to receive the digested sludge. These farms were obtained primarily through advertisements in the local newspaper.

One full-time driver is employed to haul the sludges from both of the city's plants using an over-the-road tractor-trailer unit with a tank capacity of 3,000 gallons. This is a cumbersome unit for use in the fields and frequently gets stuck necessitating towing with city-owned equipment. Maintenance costs and driver qualifications are higher for this type of unit than for operation of a normal tank truck.

The land spreading operation is conducted throughout the year. However, during adverse weather conditions, the sludges are disposed of with no further attention on old sand drying beds at an abandoned wastewater treatment plant. An estimated 1 million gallons of sludge were disposed of in this fashion during the wet weather of 1972. The environs along the route to the farms are essentially rural. However, when hauling to the abandoned treatment plant, the route is through the main business district and residential areas. Two years of operational data are summarized in Table 9-8.

TABLE 9-8

OPERATING DATA FOR XENIA, OHIO LAND SPREADING PROCEDURES

	Solids (%)	Gallons of Sludge ($\times 10^3$)	Dry Weight (tons)	Labor Costs ($)	Fuel Costs ($)	Maint. Costs ($)	Depreciation ($)	Total Cost ($)	Cost per ton ($)	Cost per 1,000 gal Sludge ($)
1971	3.5	2,295	335	7,696	236	271	2,400	10,603	31.65	4.62
1972	3.5	4,150	606	6,698	442	1,304	2,400	10,843	17.89	2.61

9.6.4 Denver, Colorado

The Metropolitan Denver Sewage Disposal District No. 1 is recycling to the soil wastewater sludge at the Lowry Bombing Range. This sludge recycling program involves the chemical treatment and vacuum filtration of the sludge at the district wastewater treatment plant in Commerce City, pumping of the resultant 16 percent solids sludge cake into 42 cubic yard dump trailers, hauling of the sludge cake to the Lowry Bombing Range, transferring of the sludge cake through a sludge transfer station to 30 cubic yard truck spreaders, spreading of the sludge cake from the truck spreaders at a rate not to exceed 25 dry tons per acre, and tilling into the soil of the sludge cake after drying.

The sludge at the Metropolitan Denver Sewage Disposal District No. 1 treatment plant consists of a mixture of raw primary sludge, anaerobically digested primary sludge from the Denver Northside treatment plant, and aerobically digested excess activated sludge. The three types of sludge are mixed in a holding tank and transferred as a mixture to the sludge process building for chemical conditioning and dewatering.

To insure against water pollution problems from the sludge application, four catch basins have been constructed near the lower end of drainage gulleys and channels at the Lowry Bombing Range. Water samples are taken from the catch basins at regular intervals when runoff water is present, and tested in the laboratory of the Metropolitan Denver Sewage Disposal District No. 1. Up to the present time no significant amounts of nitrates, nitrites, fecal coliform, and other possible problem elements have been discovered.

As part of the current expansion of the Metro Denver system, a pipeline to convey sludge to the land disposal site will be built. Also, it is planned to change the method of sludge application to a subsurface injection system which has been developed at Colorado State University. The cost of this method of sludge disposal has been \$45 to \$55/ton at Denver [12].

9.6.5 Chicago, Illinois

In 1966 the Metropolitan Sanitary District (MSD) of greater Chicago started a program to return all of the digested wastewater sludge to the soil as a liquid fertilizer and soil conditioner.

The District's three main wastewater treatment plants have a designed capacity of 1.3 bgd and produce about 1,000 tons of digested sludge solids per day. Four systems of handling sludge were in use: activated sludge was heat dried and sold for fertilizer; digested sludge was stored in lagoons; wet air oxidation was used to reduce the organic solids to ash before the sludge was stored in lagoons; Imhoff tank sludge was dewatered on drying beds before being hauled to a dump. None of these sludge handling and disposal methods were satisfactory as a permanent method.

The land reclamation project plan was developed to achieve economical disposal and utilization of the sludge. The plan provided for irrigation of low quality farmland with digested liquid sludge. About 30,000 acres located about 50 miles from Chicago were to be purchased. Pipeline transportation and lagoon storage at the terminal were part of the recommended systems.

Local opposition prevented acquistion of the desired land, and available land needing improvement was finally located in Fulton County, Illinois, nearly 200 miles from Chicago where 10,000 acres of the strip-mined land were purchased. Seventy-one percent of the land was formerly strip-mined. Under the initial temporary plan for Fulton County, 7,500 wet tons/day (50 percent of the MSD's daily sludge production) sludge is being barged down the Illinois River to a dock from which it is pumped through 11 miles of pipeline to storage lagoons. About 10 percent of the sludge being barged to Fulton County is pumped from the storage lagoons through irrigation pipe and sprayed on terraced spoil banks during 8 months of the year. Runoff from the sludge-amended land is contained in reservoirs until the water quality meets stream standards before it is discharged.

By July, 1974, 2,000 acres of the 10,000 owned were planned to be ready for sludge application — 1,100 acres of which were formerly strip-mined land and 900 acres row cropland [13]. The cost to prepare the former strip-mined land has been $2,000 to $3,000 per acre. Corn is now being grown on 900 acres. About 750 acres of the site have been established as a recreational area.

Sludge is sprayed by a carriage mounted spray gun which sprays the sludge in a 300-foot arc. Crops grown include corn, soybeans, and winter wheat. Application rates to the strip-mined soil are scheduled to decrease from an initial 75 dry tons/acre/year during the first year to 20 dry tons/acre/year over a 5-year period.

The cost of barge transportation is $36.81 per ton of sludge solids, and the total cost of disposal is $71.54 per ton of digested solids [14]. Long-range land reclamation plans for

Fulton County include construction of a pipeline from Chicago. The total cost of sludge disposal is expected to $35.24 per ton of solids [14]. Of the above totals, sludge digestion comprises $9.22/ton.

Although the Chicago experience indicates the potential for converting strip-mined land to productive agricultural land, the project is continuing to meet with some citizen opposition [13]. About 25 percent of the registered voters in Fulton County have signed a petition asking if agencies outside the county should be stopped from spreading sludge on lands there. As a result, an advisory referendum on the issues has been scheduled for November, 1974. Chief citizen concerns have been odors and the potential threats of heavy metal buildup and pollution.

9.7 References

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CHAPTER 10

CASE HISTORIES–USE OF CHEMICALS IN EXCESS ACTIVATED SLUDGE PROCESSING

10.1 General Considerations

A reliable source of information for plant design is an existing plant's operating data. Accordingly this chapter reviews performance data which was compiled over several years from a number of plants.

10.2 Washington, D.C.–Blue Plains Plant

Dahl, Zelinski, and Taylor [1] have excellently described many years of process optimization work for the 253 mgd Blue Plains plant. The D.C. plant is presently a modified, high-rate, activated sludge plant and has the processes and mode of operation shown schematically in Figure 10-1. Prior to the 1959 addition of the modified activated sludge system, the plant had all the facilities shown in Figure 10-1 except for the aeration basins and final clarifiers. The rationale for the plant expansion included the supposition that the same solids handling processes could be used for treating the mixed primary and activated sludge as had been used for the primary sludge alone. The expansion did include, however, a doubling of the elutriation tank capacity as well as some expansion in total solids handling capabilities. Key assumptions were that a mixture of raw primary sludge and excess high rate activated sludge could effectively be gravity thickened and that elutriation of the anaerobically digested mixed sludge could work well. Hindsight showed gravity thickening to be partially satisfactory, while elutriation was not.

A short time after starting operation of the modified secondary plant; the high solids content of the thickener overflow and more particularly, the elutriate, caused recycle of a large load of fine solids. The primary basins and aeration systems could not accommodate these fine solids and hydrated activated sludge particles. They built-up in the system and caused very poor capture in the primary sedimentation basin and gradually saturated the entire plant causing septicity and flotation of sludge blankets. Upsets in the various solids liquid separation procedures eventually caused essentially inoperable conditions. A temporary solution was obtained by venting the elutriate to the plant effluent, since the elutriate represented the largest recycle stream in the plant. This, however, was not a satisfactory long-range solution, because it lowered plant efficiency. In effect, the elutriate contributed 15 to 30 tons per day of BOD to the Potomac River. Early in 1965, a high molecular weight cationic polyelectrolyte was applied to the elutriation basin's feed in the manner shown in Figure 10-2. This process modification dramatically cleaned up the elutriate and offered a way to solve the elutriate problem. However, the use of chemicals in elutriation was only a trial process change and diversion of effort to other

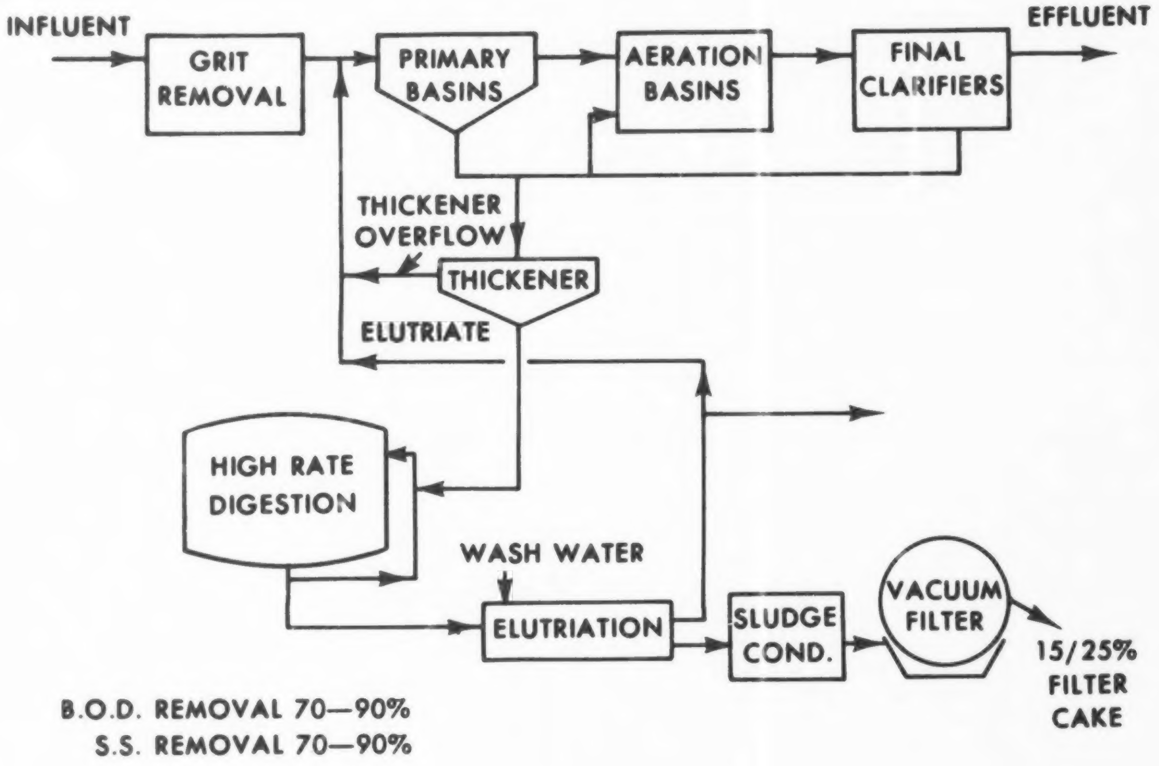
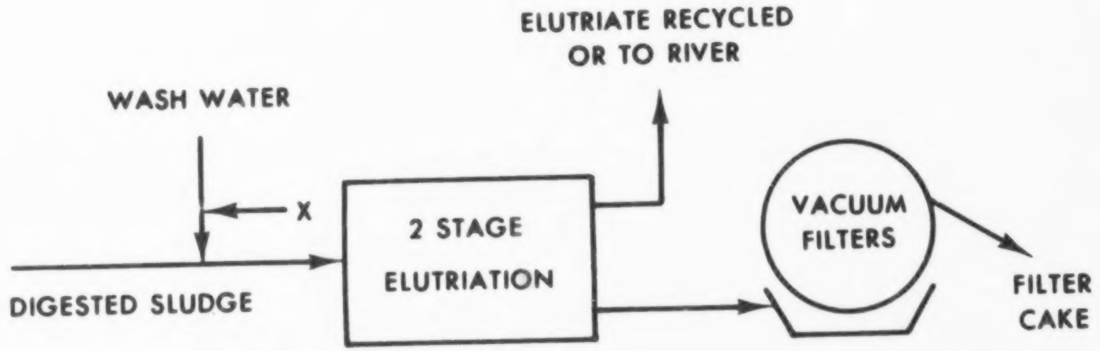

INFLUENT
GRIT REMOVAL
PRIMARY BASINS
AERATION BASINS
FINAL CLARIFIERS
EFFLUENT
THICKENER OVERFLOW
THICKENER
ELUTRIATE
HIGH RATE DIGESTION
WASH WATER
ELUTRIATION
SLUDGE COND.
VACUUM FILTER
15/25% FILTER CAKE
B.O.D. REMOVAL 70—90%
S.S. REMOVAL 70—90%

FIGURE 10-1. District of Columbia, plant flow diagram.


ELUTRIATE RECYCLED OR TO RIVER
WASH WATER
X
2 STAGE ELUTRIATION
VACUUM FILTERS
DIGESTED SLUDGE
FILTER CAKE

X = CATIONIC POLYELECTROLYTE APPLICATION POINT

FIGURE 10-2. District of Columbia's elutriation and filtration system.

plant processing studies considered potentially more efficacious delayed full realization of the benefit of application of chemicals to elutriation. During the period 1968 to 1969, the focus of plant studies was again on the elutriation and filtration systems. By the application of a cationic polyelectrolyte to the elutriation system and through intensive study of the elutriation and filtration processes, it was found possible to achieve 90 percent solids capture through the elutriation and filtration systems. The achievement of 90 percent capture produced an elutriate that could be recirculated to the head of the primary basins and not cause any upset in plant operations. Although the use of a cationic polymer to flocculate the digested mixture of primary and activated sludges affected good solids capture, settling rates, and compaction, considerable benefit was also obtained in operational control. This was brought about by the operating personnel who devised improved methods for monitoring the performance of the elutriation basins and the vacuum filters.

Shortly after the modified activated sludge process began operation and during the venting of elutriate, it was only necessary to remove about 45 tons per day of sludge solids at the filters. Ferric chloride was used for sludge conditioning and filter production rates were about at the design figure of 3 lb/hr/ft^2. Chemical costs on occasion, with ferric chloride, were as high as $13.50 per ton. Immediately after starting chemical treatment of the elutriation basins to effectively capture 90 percent of the solids, the amount of solids which had to be removed by the filters increased dramatically. This was because the plant had effectively become saturated with fine solids, resulting in a temporarily higher than normal sludge removal rate at the filters and subsequently higher chemical and operating costs than normal. After prolonged use of the polymer in elutriation, a new plant equilibrium was established and the removal rate and conditioner demand decreased. Table 10-1 shows the effect of chemical addition to the elutriation process on solids capture and chemical requirements for good vacuum filtration of sludge.

TABLE 10-1

DISTRICT OF COLUMBIA'S SLUDGE REMOVAL PRACTICES AND COSTS

	Tons/Day Removed	Chemical Cost ($/ton)	
		Elutriation	Filtration
Elutriate to River	45	—	13.50
Post Elutriate Recycle Period (Polymer in Elutriation)	80	4.68	7.42
After Prolonged Polymer Use in Elutriation	70	Total = 9.75	

Figure 10-3 shows the excellent dewatering operation at the Washington, D.C. plant. The filter yield obtained with the new mode of sludge processing at D.C. is over 4 lb/hr/ft^2, which is 33 percent above the design rate. Previous mention was made of a poor quality thickener overflow stream. The use of flocculation was limitedly investigated on a laboratory scale, and it was determined by a cost/benefit analysis that the use of flocculants in the thickener was not advisable under prevailing operating conditions.

During the current plant expansion, separate conditioning boxes and feed systems were provided for each filter. This was not originally the case but is essential for smooth operation of the vacuum filters. Experience gained at many plants has indicated that without separate conditioning and feed systems, all filters do not receive optimumly conditioned sludge.

An interesting facet of the Blue Plains plant is the fact that flash dryer type incinerators were installed at the time of expansion to secondary treatment. These incinerators have, however, remained inoperative since installation and filter cake has been disposed of on land. Land disposal in this instance was much less expensive than incineration and ash disposal. Presently the Blue Plains plant is again being expanded in capacity and upgraded to meet Potomac Estuary Standards for phosphorus and nitrogen removal. Since additional sludge will be created by the new processes, present disposal methods will require reevaluation. Figure 10-4 shows the essential features of a proposed sludge processing system. Since incineration may be practiced, digestion has been eliminated. Separate gravity thickening of primary sludge and flotation thickening of excess activated sludge have been provided, and the thickened sludges would be combined and dewatered just prior to incineration.

10.3 St. Helens–United Kingdom

The solids handling system at the St. Helens plant is shown in Figure 10-5. Excess activated sludge is recirculated to the primary basin where it is mixed with primary sludge. The combined primary and activated sludges are then anaerobically digested, elutriated and vacuum filtered. During the period 1965 to 1968, the amount of excess activated sludge to be handled by the plant increased significantly. This increase in the quantity of EAS was attributed to an increasing degree of treatment, low retention time in the aeration system, and recycling of a high solids elutriate stream. Since the sludge going to the vacuum filter contained a very large amount of EAS, the filter yield deteriorated greatly [2,3]. The four-year trend in the solids content of the elutriated digested sludge fed to the filters is shown in Table 10-2.

Figure 10-6 shows that the variation in filter yield coincided with the variation in the solids content of the elutriated sludge feed to the filters. Not unexpectedly, as the sludge solids increased, so did the filter yield. Figure 10-7 further shows that the solids content of the elutriated sludge fed to the filters varied with quantity of excess activated sludge produced. As the amount of excess activated sludge increased, the degree of solids concentration achieved in elutriation decreased.

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FIGURE 10-3. Vacuum filter operation at District of Columbia.

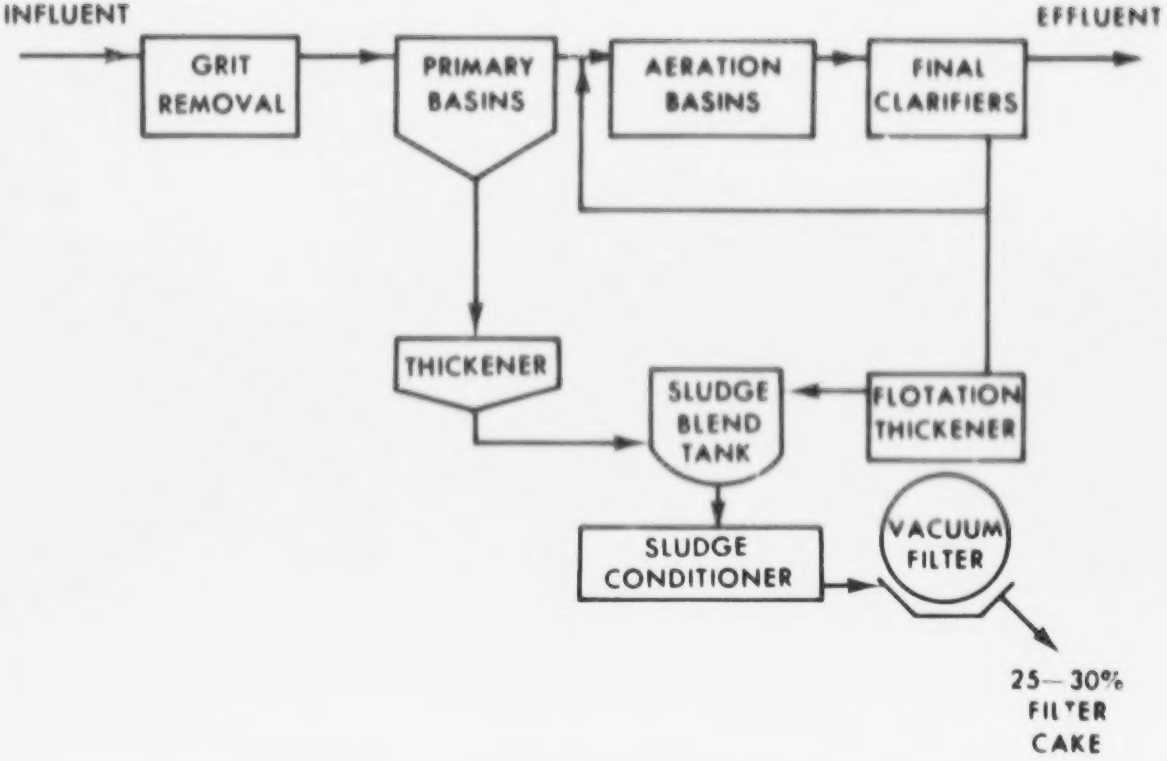
INFLUENT
GRIT REMOVAL
PRIMARY BASINS
AERATION BASINS
FINAL CLARIFIERS
EFFLUENT
THICKENER
SLUDGE BLEND TANK
FLOTATION THICKENER
SLUDGE CONDITIONER
VACUUM FILTER
25— 30% FILTER CAKE

FIGURE 10-4. New Blue Plains sludge processing system.

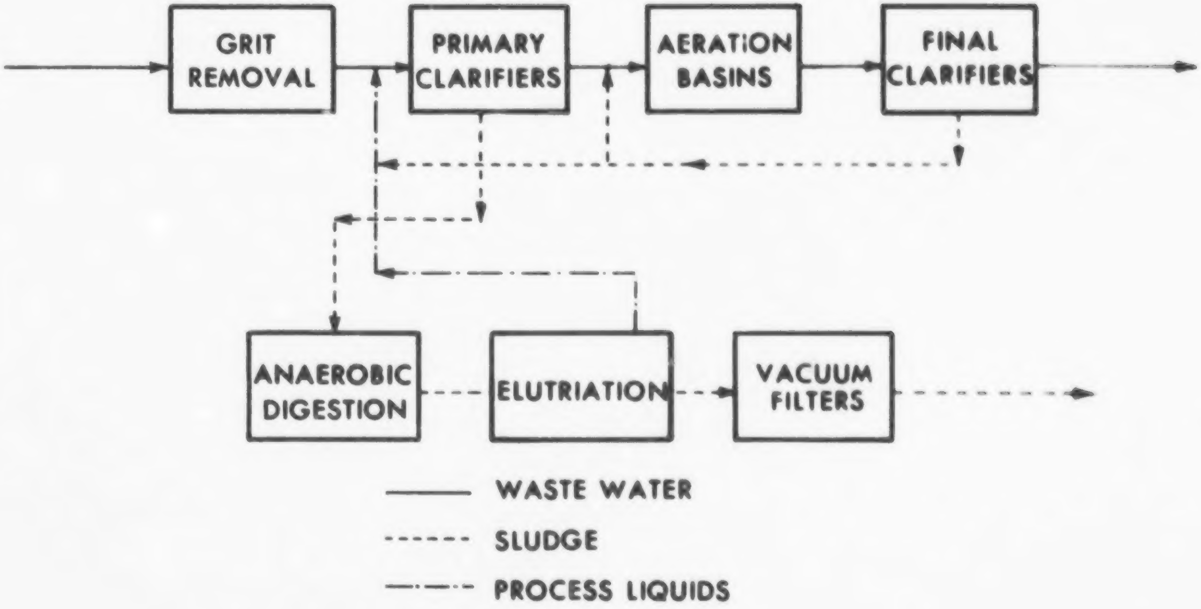
GRIT REMOVAL
PRIMARY CLARIFIERS
AERATION BASINS
FINAL CLARIFIERS
ANAEROBIC DIGESTION
ELUTRIATION
VACUUM FILTERS
WASTE WATER
SLUDGE
PROCESS LIQUIDS

FIGURE 10-5. Solids handling at the Parr Works, St. Helens.

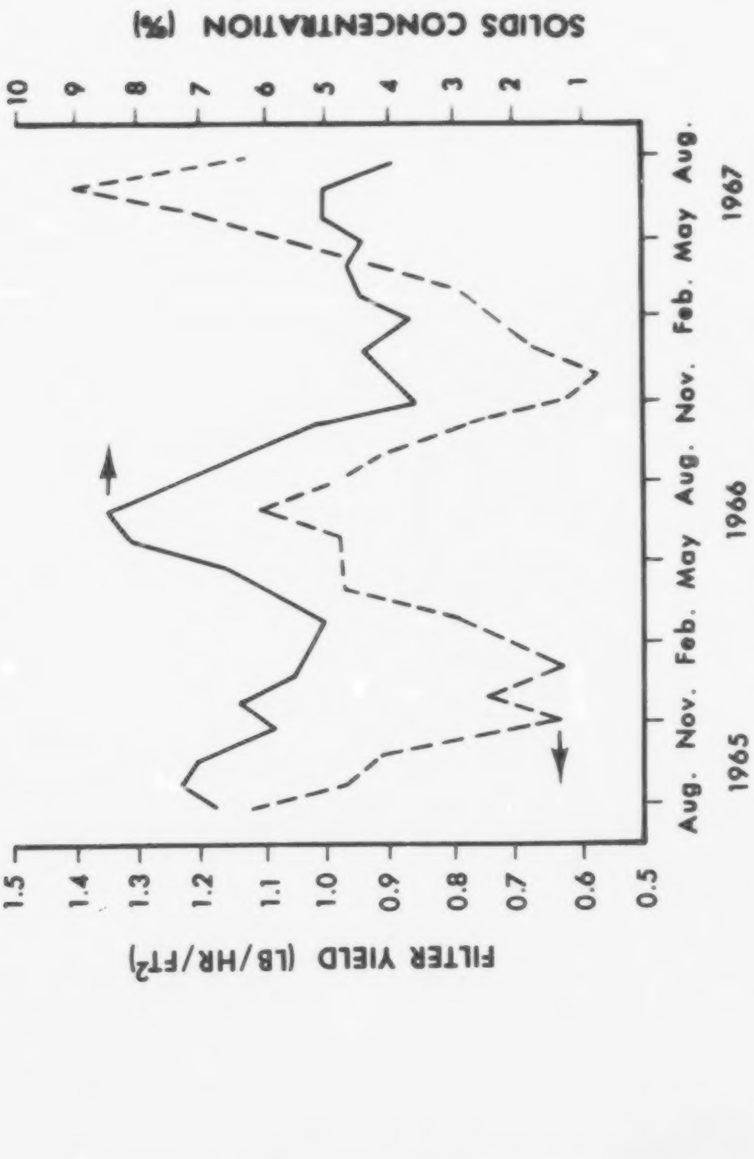

FIGURE 10-6. Variation of filter yield and percent solids with time at St. Helens.

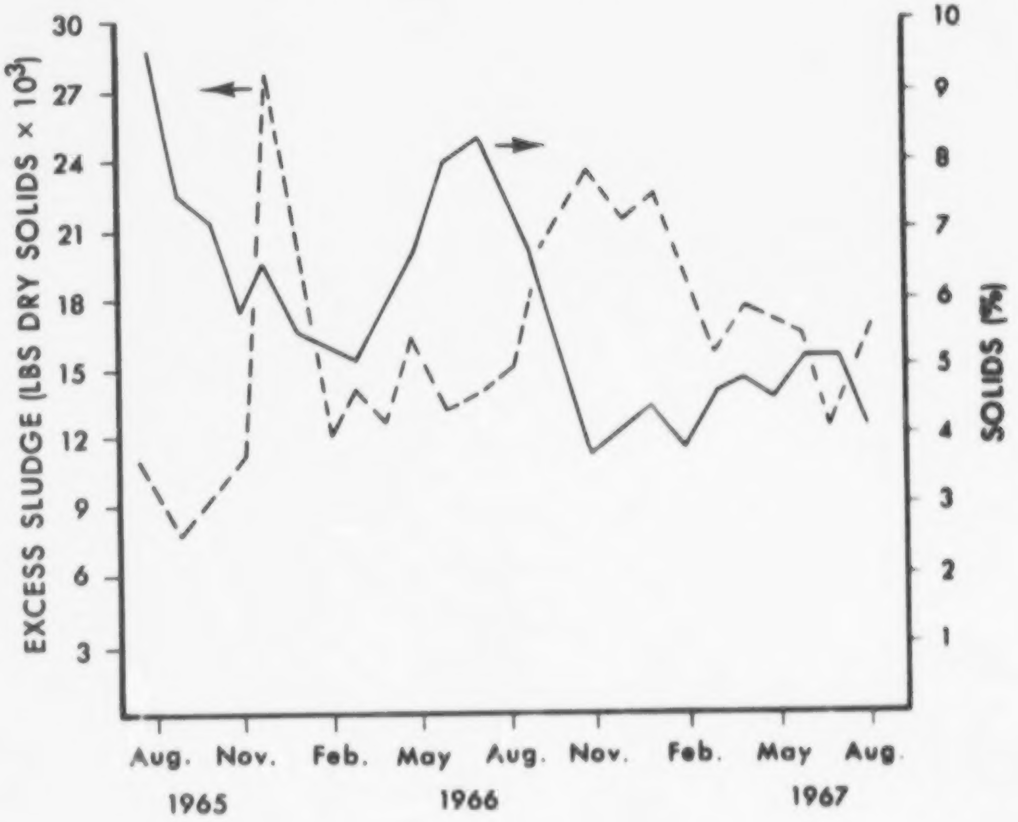

FIGURE 10-7. Quantity of EAS and solids concentration in the EAS as a function of time.

TABLE 10-2

VARIATION OF PERCENT SOLIDS IN ELUTRIATED SLUDGE AT ST. HELENS

1965	1966	1967	1968
7.16	5.68	4.08	2.08

The low aeration detention time at St. Helens is partially responsible for a greater production of excess activated sludge. Table 10-3 compares the St. Helens plant with the Hogsmill Valley plant [4]. The latter plant has a considerably longer detention time and permits some endogenous respiration to occur and partially destroy the EAS solids. Composition of the influent wastes, particularly their inert content can also affect these figures.

TABLE 10-3

EFFECT OF AERATION ON EXCESS ACTIVATED SLUDGE PRODUCTION

Plant	Aeration Retention Time (hr)	EAS lb/lb BOD Removed
St. Helens	4.68	1.44
Hogsmill Valley	9.9	0.74

Some success in improving filter yield has been obtained by using intermittent storage of sludge in a lagoon to facilitate smooth operation of the elutriation system. In addition to use of the lagoon to obtain a thicker sludge, procedural changes and improvements in sludge conditioning equipment and dosage control have resulted in a cleaner elutriate and a somewhat higher filter yield. Attempts to demonstrate effectiveness of polymer flocculation in the elutriation system at St. Helens have been unsuccessful to date [2,3].

10.4 Metropolitan Toronto Main Plant

The Metro Toronto main plant is a step aeration activated sludge plant featuring two-stage anaerobic digestion of the mixed primary and activated sludge, elutriation, vacuum filtration with coil filters, and multiple hearth incineration. The plant's flow diagram is shown in Figure 10-8. Excess activated sludge is recirculated to the head of the plant. Considerable recirculation options exist for excess activated sludge, digester supernatant, elutriate, and filtrate, however, which are not accurately reflected in the figure.

Both the hydraulic and organic loading on the plant increased in the period 1967 to 1970, and the plant was expanded in the interim from 1969 to 1971 by installing additional aeration basins and final clarifiers [5]. As might have been expected, the additional pollutant load on the plant and the expansion of the activated sludge process increased the quantity of sludge to be processed. Since there was now a much larger quantity of EAS to be handled, as was the situation at Washington, D.C. and St. Helens, a gradual decrease occurred in the solids concentration of the elutriated digested mixed sludge fed to the filters. This is shown in Figure 10-9. The situation did not become critical until 1970, when the average monthly solids concentration for the elutriated sludge was consistently below 4 percent. The filters were then unable to remove the sludge solids from the system by their usual mode of operation. During August 1970, the solids content of the elutriated sludge being fed to the filters even dropped to less than 3 percent and was nonuniform. The nonuniformity and low solids caused low filter yields and wet filter cake. Figure 10-10 clearly shows the long-term trend for the solids concentration of the raw sludge, which was fed to the digesters. The solids concentration fell from a level of 6 percent to 4 percent and below during 1970. This trend is also a reflection of the increasing amounts of excess activated sludge being handled at this time. To maintain reasonable plant operation, solids processing data indicated the four Komline-Sanderson coil filters should process at least 3.7 lb D.S./hr/ft^2. Further, in 1971 the filters would need to process at least 4.4 lb D.S./hr/ft^2. This information is shown in Table 10-4.

TABLE 10-4

METRO TORONTO'S SLUDGE REMOVAL NEEDS

	Operable 1970	Preferred 1970	Required 1971
Dry tons/mo	2,000	2,500	3,000
Filter Productivity (lb/hr/ft^2)	3.0	3.7	4.4

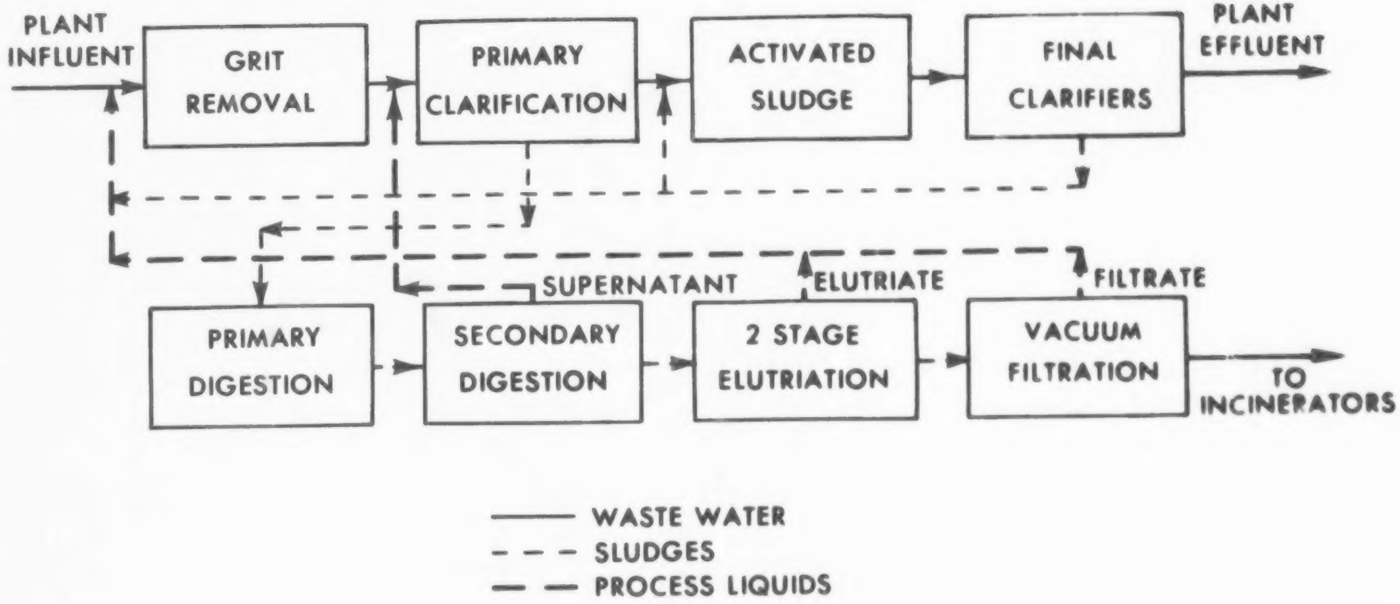
PLANT INFLUENT
GRIT REMOVAL
PRIMARY CLARIFICATION
ACTIVATED SLUDGE
FINAL CLARIFIERS
PLANT EFFLUENT
SUPERNATANT
ELUTRIATE
FILTRATE
PRIMARY DIGESTION
SECONDARY DIGESTION
2 STAGE ELUTRIATION
VACUUM FILTRATION
TO INCINERATORS
WASTE WATER
SLUDGES
PROCESS LIQUIDS

FIGURE 10-8. Metro Toronto's plant flow diagram

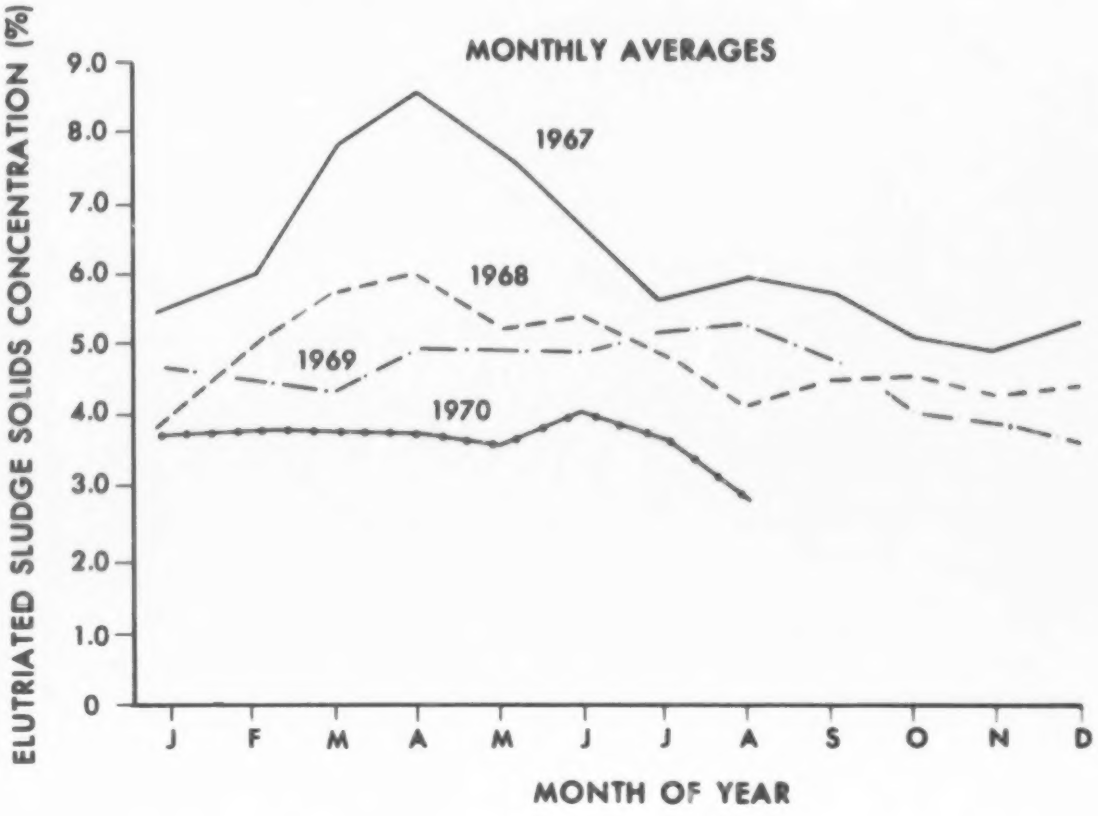

FIGURE 10-9. Variation of percent solids in elutriated sludge at Metro Toronto for period, 1967-1970.

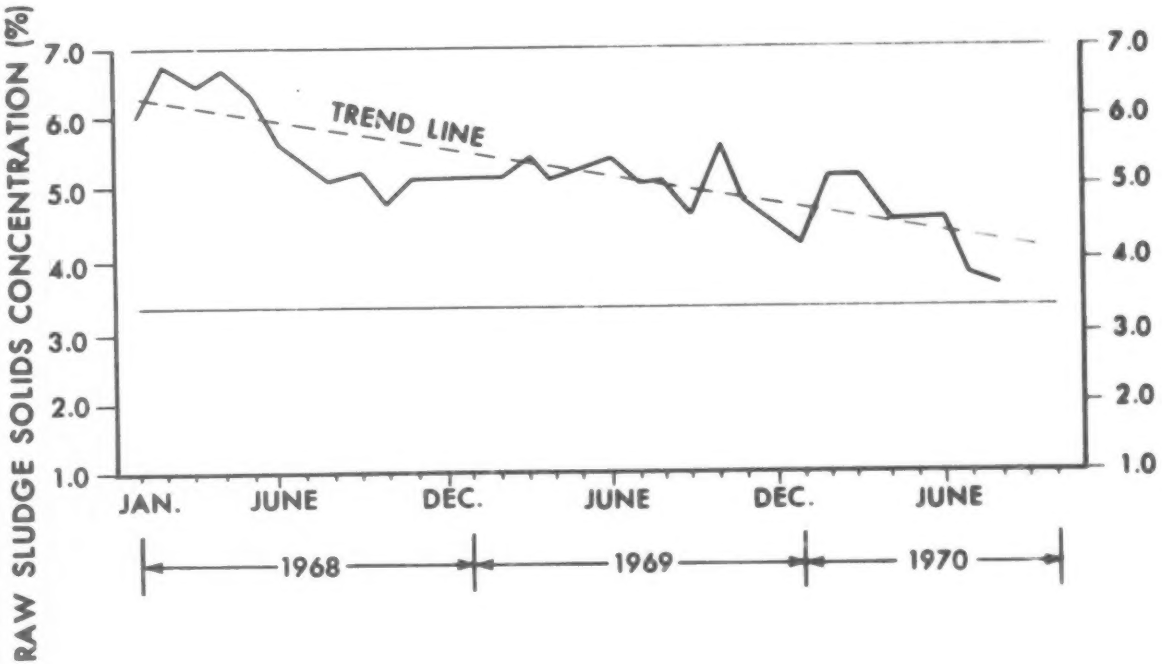

FIGURE 10-10. Metro Toronto's raw sludge solids concentration from 1967 to 1971.

At this juncture plant personnel decided to make a cost benefit determination for the use of small quantities of polymer in the elutriation tanks. During October and November 1970, between one and two pounds of polymer were added to the elutriation basins per ton of dry solids feed. Results are shown in Table 10-5. Concurrently, small and decreasing amounts of ferric chloride were used with the polymer at the vacuum filters. Immediate and dramatic improvements were noted in the concentration and uniformity of the elutriated sludge solids concentration, which resulted in smooth filter operation and yields in excess of 4.3 lb/hr/ft^2. Total solids capture in the elutriation and filtration processes was controlled at the 90 percent level.

TABLE 10-5

ELUTRIATION/FILTRATION RESULTS FOR OCTOBER/NOVEMBER AT METRO TORONTO

1970 Period	Polymer Used (lb/ton)		Sludge Solids	Yield	Cake Solids
	Elut.	Filt.	(%)	(lb/hr/ft^2)	(%)
October	1.26	7.77	3.6	4.7	16
November	1.75	8.20	4.1	4.3	16

Since the polymer contract year ended during January 1971, a number of polymers and various modes of elutriation system operation were evaluated. Some results are shown in Table 10-6. It was possible to vary the solids concentration of the elutriated sludge by varying the loading, wash water ratio, and polymer dosage to the elutriation basins. No ferric chloride was necessary to prepare sludge for filtration. As the results in Table 10-6 show, filter yields of 4.7 to 5.8 lb/hr/ft^2 were obtained at different loadings, wash water ratios, and polymer dosages. In general the uniformity and ease of operation of the filtration process were dramatically improved with the new mode of operation. Figure 10-11 shows the excellent cake discharge obtained and the thickness of the cake.

In effect, the sludge processing improvements just described cleared up one of the two major recirculation streams inherent in the Toronto design. It did not, however, eliminate the large recirculation of solids and soluble BOD from the two-stage digestion system. This recirculation load has a deleterious effect on the amount of excess activated sludge produced and, hence, the relative quality of mixed sludge fed to the elutriation, filtration, and incineration systems. It is possible that separate flotation thickening and aerobic

TABLE 10-6

ELUTRIATION AND FILTRATION RESULTS DURING 1971 AT METRO TORONTO

Elut. Flow (mgd)		Polymer (lb/ton)		Elutriate S.S. (mg/l)		Sludge Solids (%)	Vacuum Filter Performance	
							Yield (lb/hr/ft²)	Cake Solids Concentration (%)
Wash Water	Digested Sludge	Elut.	Filt.	1st	2nd			
1.0	0.6	1.94	10.96	120	18	6.1	4.7	16.0
3.5	1.4	0.62	9.34	6,250	208	3.5	5.8	15.4

digestion of the activated sludge, followed by mixing of aerobically digested activated sludge with either raw or anaerobically digested primary sludge would materially improve plant costs. This proposed process improvement would provide a readily controllable method of thickening the excess activated sludge and reduce the amount of excess activated sludge to be processed. It also would keep the activated sludge out of the anaerobic digestion system which operates more efficiently with raw primary sludge and produces a better quality supernatant. Other benefits of the process improvement would be a reduced chemical demand for the sludge feed to the vacuum filters and an improved cake solids concentration in the feed to the incinerators.

10.5 Richmond, California

From 1967 to 1969 the city of Richmond, California, modified its wastewater treatment plant to provide secondary treatment by the activated sludge process. Figure 10-12 shows the plant flow diagram. The Richmond plant has a maximum hydraulic capacity of 40 mgd with an average daily flow of 9 mgd for the period of 1969 to 1970. Flotation thickeners were included in the plant's design to thicken EAS. Digester supernatant and filtrate are recycled to the elutriation system. The mixed digested elutriated sludge is vacuum filtered and either incinerated or landfilled [6]. It was speculated during design that the inclusion of flotation thickening of the activated sludge was a positive step, but that mixing of the primary and secondary sludges early in the solids handling process might give problems in subsequent solid liquid separation processes such as elutriation and dewatering. Shortly after the activated sludge system was put into operation, the same problems that occurred at other plants occurred at Richmond. Solids accumulated

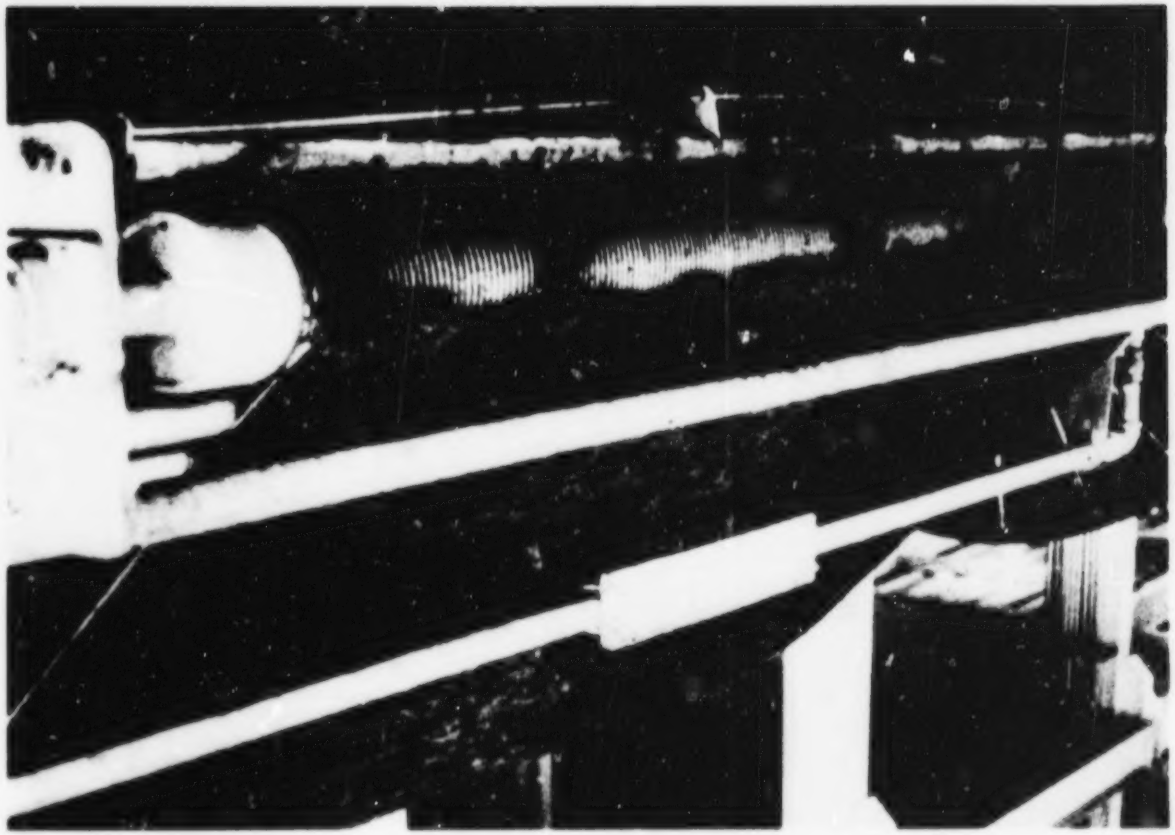

FIGURE 10-11. A view of filters at Metro Toronto.

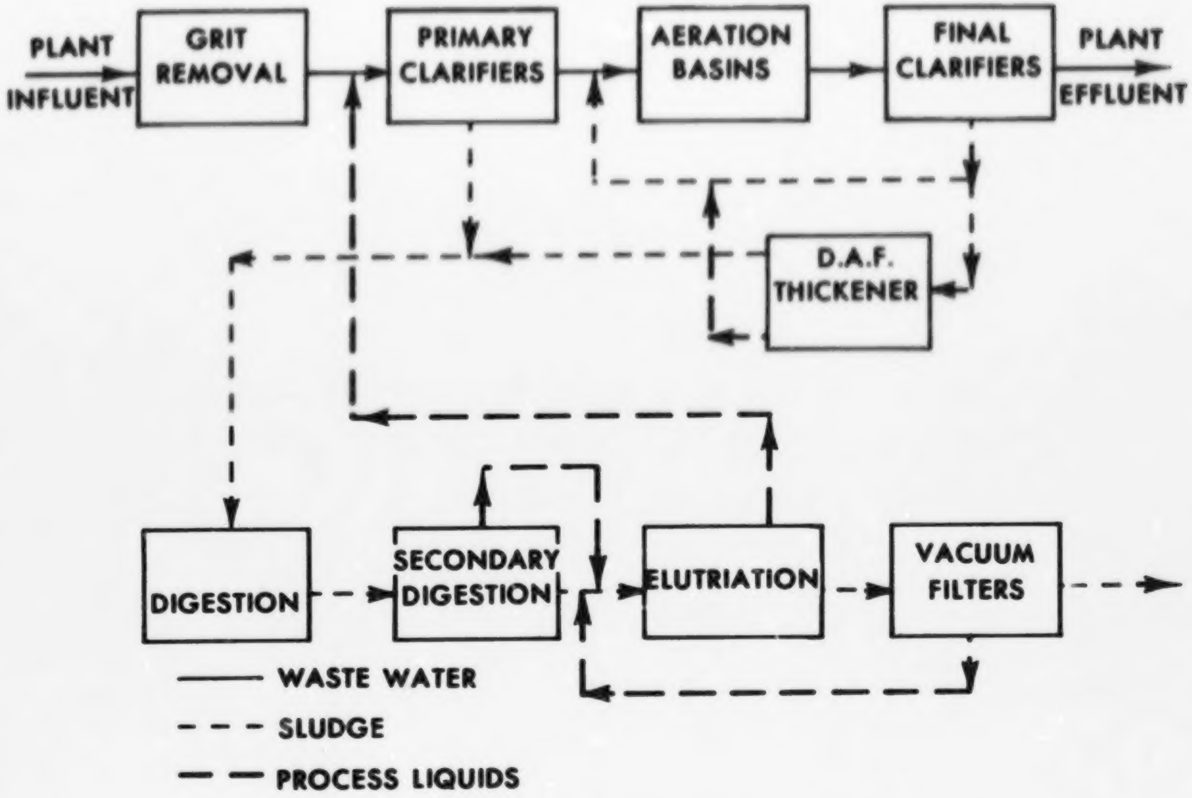
PLANT
INFLUENT
GRIT REMOVAL
PRIMARY CLARIFIERS
AERATION BASINS
FINAL CLARIFIERS
PLANT
EFFLUENT
D.A.F. THICKENER
DIGESTION
SECONDARY DIGESTION
ELUTRIATION
VACUUM FILTERS
WASTE WATER
SLUDGE
PROCESS LIQUIDS

FIGURE 10-12. Richmond, California's plant flow diagram.

in the plant because of poor capture and compaction in elutriation, and low vacuum filter yields and high operating costs were experienced. After much process improvement work, plant personnel adopted regular use of a cationic high molecular weight polymer in the elutriation basins to alleviate their solids handling problems. The cost and efficiency of the dewatering operation was dramatically improved by the use of the polymer in elutriation and this is shown by the data in Table 10-7. Additional data on the filtration and elutriation operations are presented in Table 10-8. Excellent capture and compaction were achieved in elutriation. About 70 percent of the conditioning chemical cost for sludge processing was due to the ferric chloride and lime dosages required for good vacuum filtration.

TABLE 10-7

VACUUM FILTRATION RESULTS FOR RICHMOND, CALIFORNIA

		Mixed Sludges	
	Primary Sludge	No Polymer	Polymer in Elutriation
Yield (lb/hr/ft^2)	7-9	1-2	5-7
Conditioner Cost ($/ton)	3.80-4.00	25-30	11-14
Cake Solids (%)	29-31	16-18	20-22

TABLE 10-8

RICHMOND, CALIFORNIA–ELUTRIATION AND FILTRATION OPERATIONS' DATA

Process	Digest Sludge Solids Conc. (%)	Elutriate Sludge Solids Conc. (%)	Quantity of Polymer Added (lb/ton)	Elutriate Solids Conc. (mg/l)
Elutriation	3.85	7.8	2.12	450
	$FeCl_3$ Dosage ($/ton)	Lime Dosage ($/ton)	Filter Cake Solids Conc. (%)	
Filtration	3.00	4.85	20.8	–

Still further economies might be attained if the dewatering equipment at Richmond had an adequate sludge cake release mechanism. The belt filters now used have essentially no release mechanism, and this necessitates overconditioning in terms of ferric chloride and lime dosage to obtain cake discharge. This situation is clearly shown in Figure 10-13, where the drying cracks in the cake on the belt are evident. The cracked cake appears a few inches above the point where the drum comes out of the vat, and this is an abnormal condition. Since these conditions were reported, Richmond personnel have partially alleviated the release problem by use of polypropylene type filter cloth. They have also eliminated the use of lime.

10.6 Fairfax County, Virginia–Westgate Plant

Robson et al. [7] have detailed the results of conversion of an overloaded intermediate treatment level plant into an oxygen activated sludge plant which removes 90 percent of the BOD_5 in the incoming wastewater. Considerable data on the processing of oxygen activated sludges have been generated at this plant. As depicted in Figure 10-14, the wastewater was originally treated at the Westgate plant by comminution; sedimentation, aeration of the mixed liquor, and secondary clarification in one long baffled tank; and chlorination. The waste sludge was digested, vacuum filtered, and landfilled.

It is not usual practice to have the three unit processes of sedimentation, aeration, and clarification in one tank. A longitudinal sectional view of this tank appears in Figure 10-15. This plant was originally designed to provide an intermediate level of treatment for a flow of 8.0 mgd. By 1970 the plant was significantly overloaded, and for a period of time chemical precipitation was practiced to upgrade plant effluent quality. The performance of the Westgate plant over approximately a 20-year period is summarized in Table 10-9.

TABLE 10-9

WESTGATE PLANT PERFORMANCE

Period	Design Flow (mgd)	% Removal of BOD_5	Plant Process
1954	8	50+	Original
1970	12	35-40	Original
1971	12	75+	Chemical Ppt.
1971-72	12	80-90	Oxygen Activated Sludge

Reproduced from best available copy.

FIGURE 10-13. Belt filters at Richmond, California.

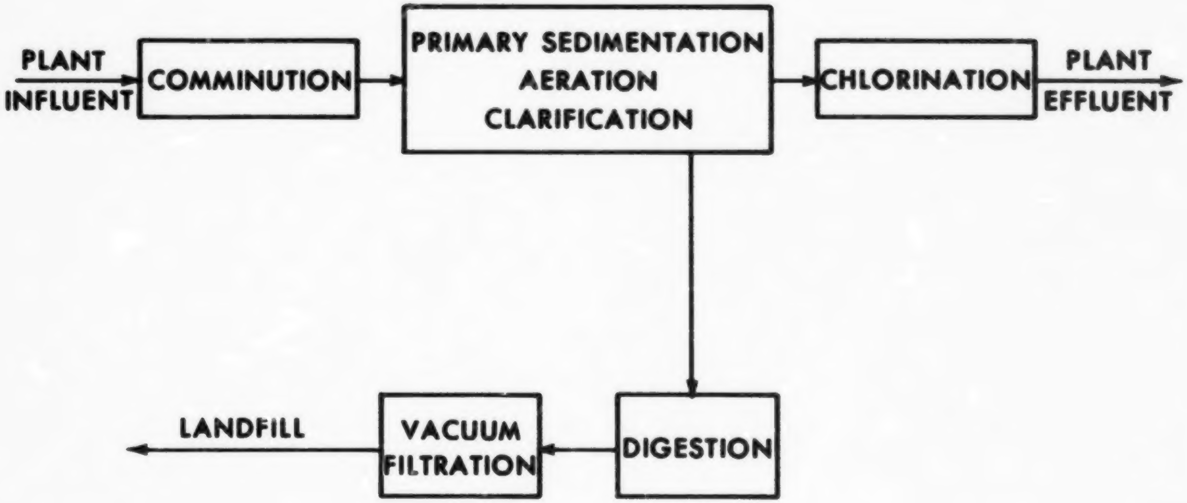
PLANT
INFLUENT
COMMINUTION
PRIMARY SEDIMENTATION
AERATION
CLARIFICATION
CHLORINATION
PLANT
EFFLUENT
DIGESTION
VACUUM
FILTRATION
LANDFILL

FIGURE 10-14. Original process flow diagram for Westgate plant.

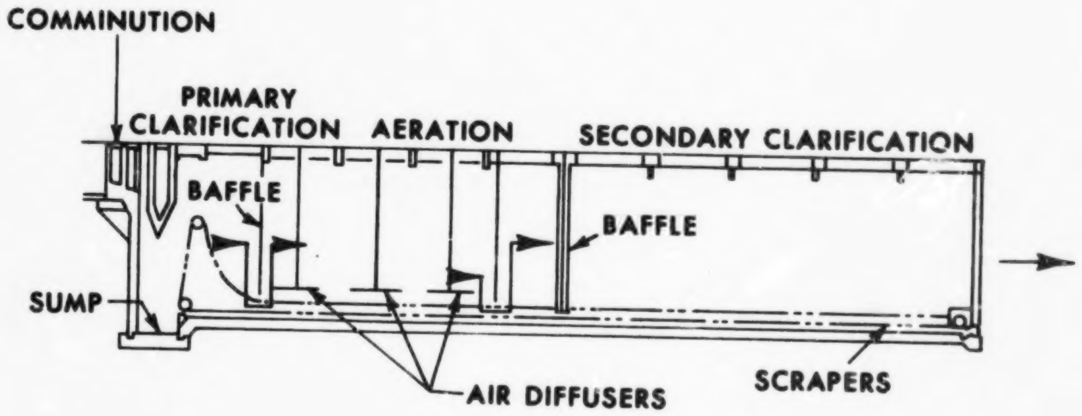
COMMINUTION
PRIMARY CLARIFICATION
AERATION
SECONDARY CLARIFICATION
BAFFLE
BAFFLE
SUMP
AIR DIFFUSERS
SCRAPERS

FIGURE 10-15. Westgate sedimentation tank.

The interim chemical treatment during 1971 was successful in upgrading the plant at a relatively high treatment cost. Although present plans are to phase out the Westgate plant within a few years, an oxygen injection system and flotation units were installed during 1971 to provide better wastewater treatment and adequate sludge processing facilities. The current Westgate plant flow diagram is depicted in Figure 10-16. Because of the temporary nature of the plant, no oxygen generation equipment was included. Rather, the plant depends on bulk oxygen supplied by a tank truck. The new wastewater treatment process has been highly successful. BOD and suspended solids removals have exceeded goals. Stable operation of the plant has become routine, and the oxygen cost has been a little lower than predicted. A summary of the actual results achieved appears in Table 10-10.

TABLE 10-10

WESTGATE OXYGEN PROCESS RESULTS

%Removal of			EAS Production	Zone Settling Velocity (ft/hr)
BOD$_5$	TSS	SVI	$\left(\frac{\text{lb VSS}}{\text{lb BOD Removed}}\right)$	
93+	90+	35-56	0.33	6.0

In addition to good wastewater treatment, it can be seen that the mixed liquor had good settling properties as illustrated by the SVI level and zone settling velocity. The amount of excess activated sludge produced was low at 0.33 lb/lb of BOD removed. Although the dissolved air flotation units functioned, it was learned during plant tests that the oxygen activated sludge settled well. This permitted bypassing the flotation units and merely gravity thickening a combination of the primary and excess activated sludges. Results at Westgate for gravity thickening of the combined sludges and vacuum filtration of the thickened sludge are shown in Table 10-11.

A small dose of flocculant is seen to be sufficient for production of a clear thickener supernatant and thickened sludge solids concentration of 6 to 8 percent. The relatively high solids concentration of the thickened sludge going to the vacuum filters results in good production rates and a resultant cake solids concentration of 22 to 28 percent. Lime and ferric chloride were used for chemical conditioning, and costs were only about $5/ton dry solids. Sludge cake is hauled to a landfill for disposal. A key point in the success of this gravity thickening process for the mixed sludge was the installation of mixers on the sludge decant tank to provide adequate blending. Figure 10-17 shows the proximity of the plant to residential areas, which shows the need for adequate odor control.

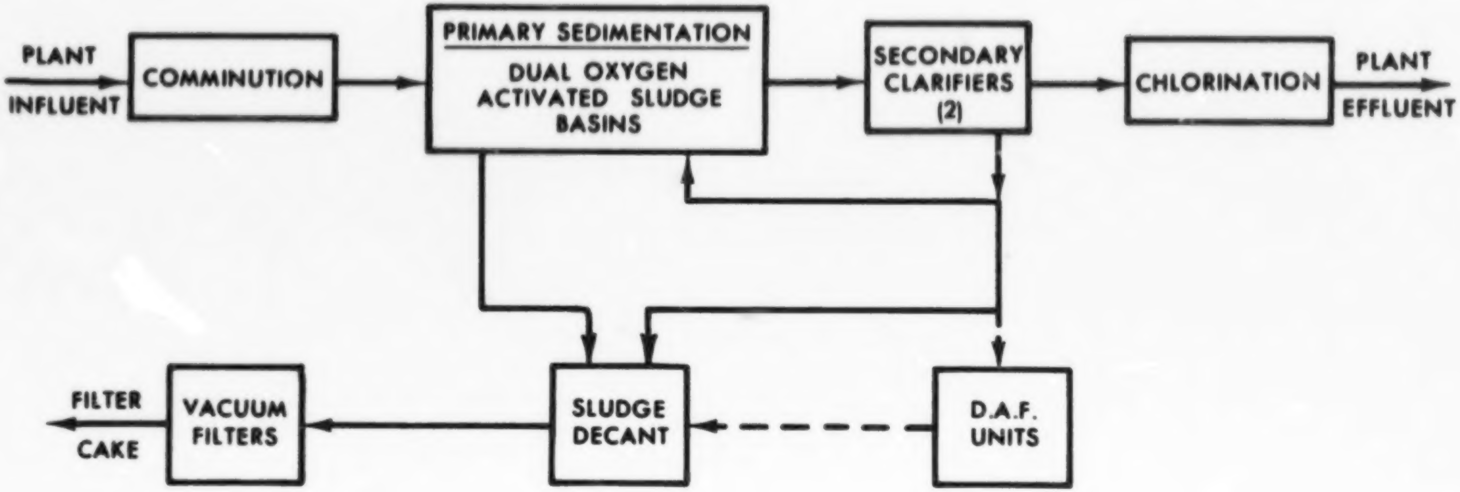
PLANT
INFLUENT
COMMINUTION
PRIMARY SEDIMENTATION
DUAL OXYGEN ACTIVATED SLUDGE BASINS
SECONDARY CLARIFIERS (2)
CHLORINATION
PLANT
EFFLUENT
FILTER
CAKE
VACUUM FILTERS
SLUDGE DECANT
D.A.F. UNITS

FIGURE 10-16. Current Westgate plant flow diagram.

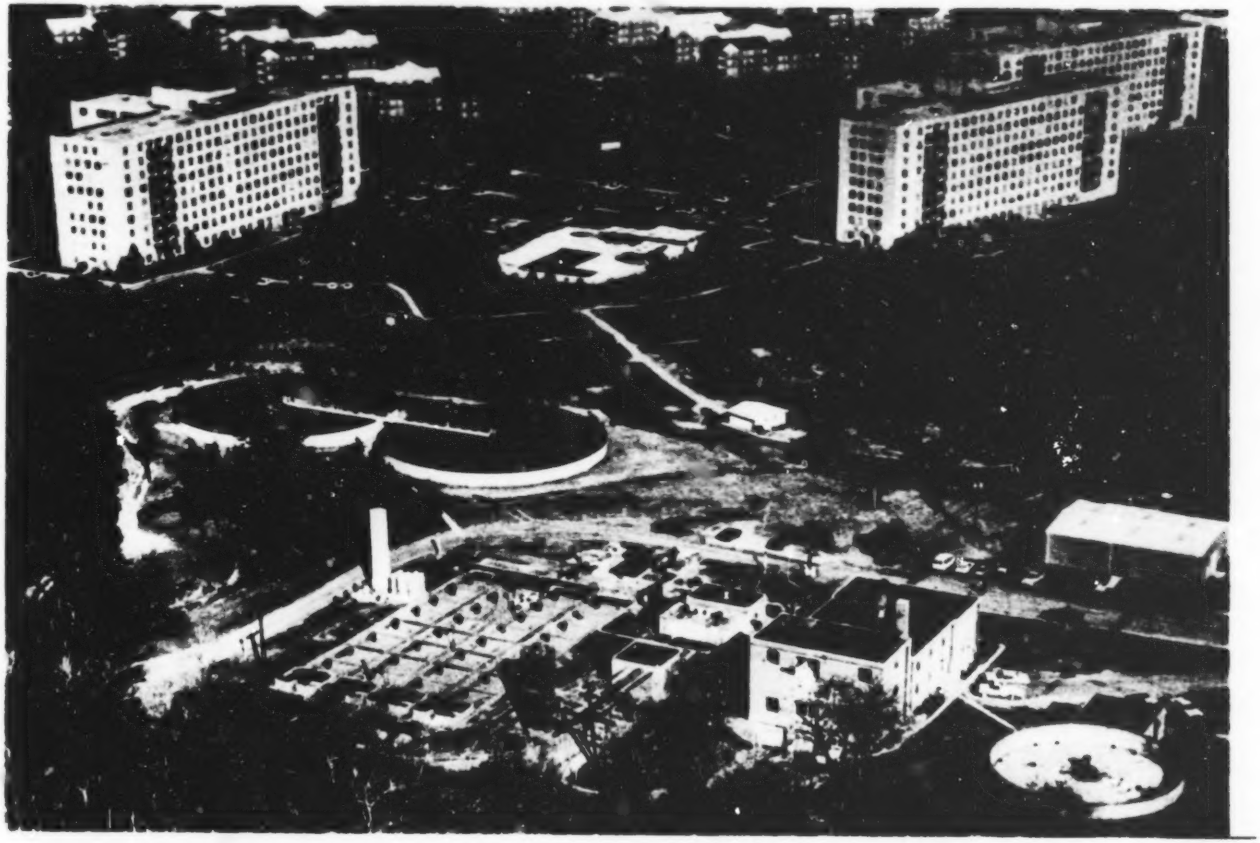

FIGURE 10-17. Fairfax County's Westgate plant.

TABLE 10-11

RESULTS FOR THICKENING AND VACUUM FILTRATION OF WESTGATE PROCESS SLUDGE

	Thickening		Vacuum Filter Performance	
Method	Polymer Dosage (lb/ton)	Thickened Sludge Solids Concentration (%)	Yield (lb/hr/ft²)	Cake Solids Concentration (%)
Gravity	3	6-8	4.0-5.0	22-28

10.7 Metropolitan Denver Sewage Disposal District No. 1

10.7.1 General Considerations

The Metro Denver plant, which commenced operation in 1966, treats wastes from the suburban Denver area as well as both settled wastewater and anaerobically digested sludge from the city of Denver's Northside plant [8]. The plant's flow diagram is shown in Figure 10-18. The Metro plant employs primary clarification, diffused air activated sludge, and final clarification. Notable factors in plant operation are the flows from the Denver Northside plant of settled wastewater into the aeration system and digested sludge into the sludge processing system. Nominal designed plant capacities are shown in Table 10-12. It should be particularly noted that the ratio of the raw and digested primary sludges to the excess activated sludge was forecast to be 0.99.

10.7.2 Sludge Processing System

The original plant design envisioned sludge processing to include dissolved air flotation thickening of the EAS; followed by dewatering of a mixture of the DAF thickened EAS, digested primary sludge, and raw primary sludge via rotary vacuum filters; and flash drying of the sludge cake prior to its use as a soil conditioner.

10.7.3 Plant Loadings Experienced

The original design had contemplated a slow rise in the process flows needing primary and secondary treatment. However, the flow magnitudes grew very rapidly, and secondary capacity was reached in 1970. Primary capacity was reached in 1972. The growth of the

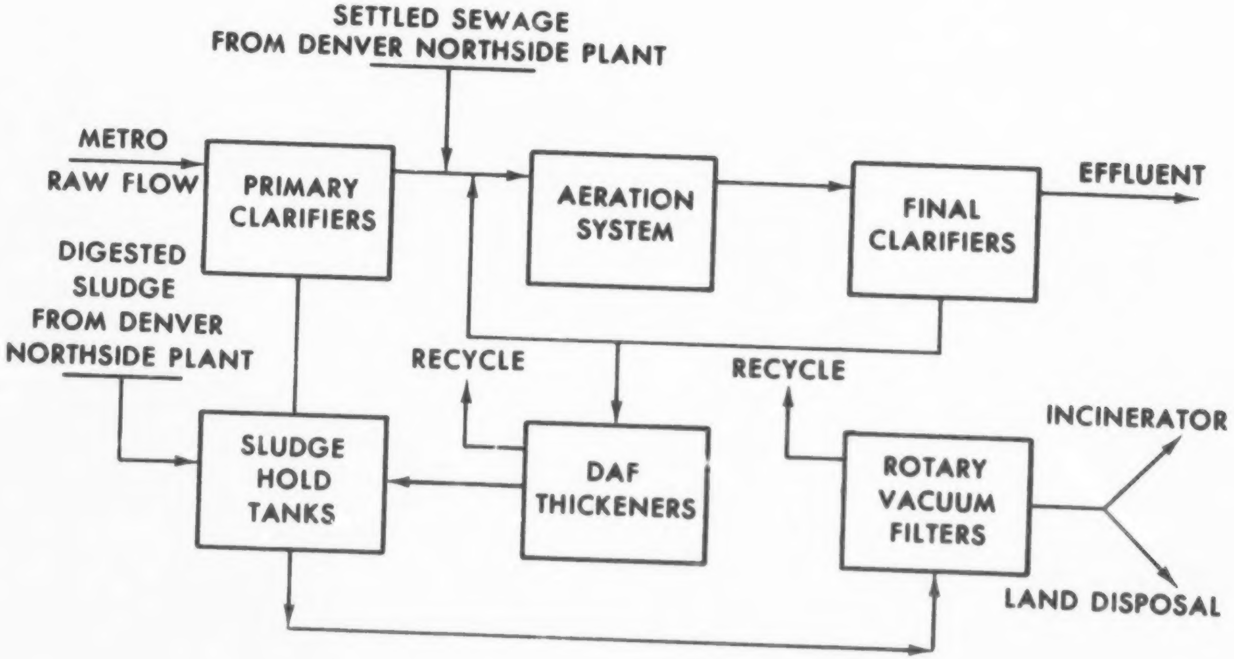
SETTLED SEWAGE
FROM DENVER NORTHSIDE PLANT
METRO
RAW FLOW
PRIMARY CLARIFIERS
AERATION SYSTEM
FINAL CLARIFIERS
EFFLUENT
DIGESTED SLUDGE FROM DENVER NORTHSIDE PLANT
RECYCLE
RECYCLE
INCINERATOR
SLUDGE HOLD TANKS
DAF THICKENERS
ROTARY VACUUM FILTERS
LAND DISPOSAL

FIGURE 10-18. Metro Denver system's flow diagram.

TABLE 10-12

METRO DENVER PLANT CAPACITIES
(Estimated 1970 Population of District: 870,000)

Processes	Average Flow (mgd)	Maximum Flow (mgd)	BOD (lb/day)
Primary Treatment	28	50	–
Secondary Treatment	98	234	166,350

Sludge Processing	Raw Primary (lb/day)	Digested Primary (lb/day)	Secondary (lb/day)
Metro Plant	37,400	–	131,000
Denver Northside	–	92,700	–

quantity of wastewater requiring primary treatment and the quantity of primary effluent requiring secondary treatment over about a nine-year period is shown in Figure 10-19.

A development which had a pronounced effect on the sludge processing operation was that only about one-third of the anticipated Denver Northside digested primary sludge solids resulted. This is reflected in the figures below which project 1975 loadings from the plant's operating history.

	Denver Northside Digested	Metro Primary	Metro Excess Activated	Total
Total dry sludge solids (lb/yr), 1975	31,000	33,000	132,000	196,000

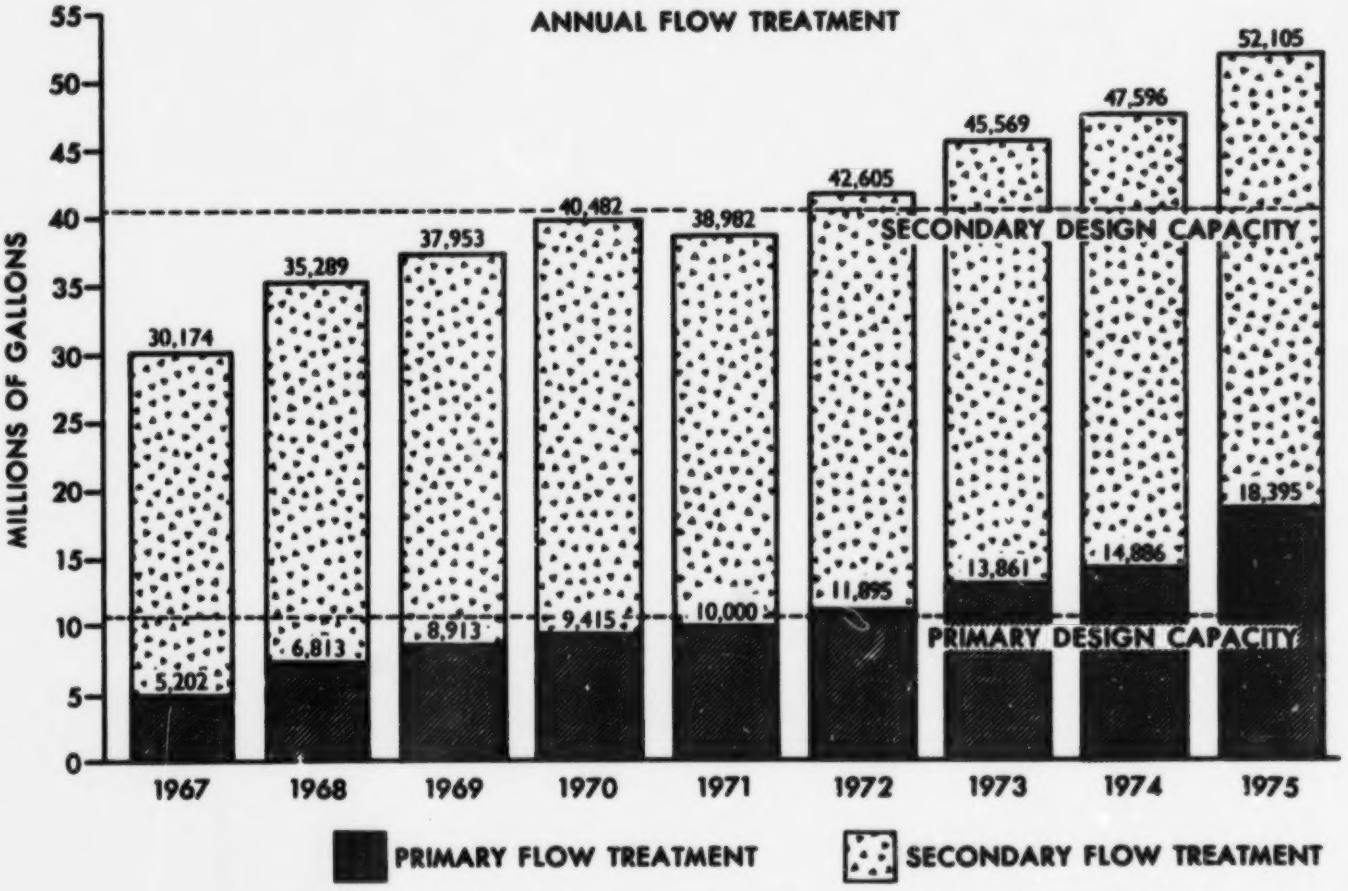

FIGURE 10-19. Changes in Metro Denver's annual plant flows from 1967 to 1975 [9].

These figures show that the ratio of raw and digested primary to excess activated sludge will be 0.485 in 1975 instead of a design consideration of a ratio of 0.99.

10.7.4 Sludge Processing Results–1967 to 1970

Considerable difficulty was encountered in sludge processing. Consistency of the sludge going to the vacuum filters was hampered by difficulty in blending the sludges in the hold tanks [9]. The higher than expected EAS to raw and digested primary sludge ratio caused severe problems with the dewatering and drying or incineration operations. Design capacities were generally not achieved.

Unit process costs for Metro Denver are shown in Figure 10-20 and were higher than anticipated. The cost picture worsened as the quantity of EAS increased. Also the chemical conditioning costs doubled after sludge was subjected to pipeline transport from one plant to another.

Compilation of data over several years resulted in the correlation shown in Figure 10-21. The high costs and difficulty encountered in attempting to process the ever increasing amounts of sludge produced led the Metro Denver staff to test and use alternate processing systems. To alleviate the sludge disposal problem, land disposal of dewatered filter cake was practiced in place of drying or incineration.

10.7.5 Modified Denver System and Results

Various operational refinements of a general nature contributed to improvements in operability and costs. In 1970 8.0 mgd of aeration capacity was adapted for aerobic digestion of the excess activated sludge. This modification proved to be a major improvement by diminishing the amount of EAS to be processed. Figure 10-22 shows the flow diagram of the modified Metro Denver system. The reduction in sludge processing costs from $60/ton to $40/ton for 1971 to 1972 was directly attributed to the partial destruction of the organic portion of the activated sludge by aerobic digestion [9]. The total Metro Denver treatment cost delineation is shown in Figure 10-23. From 1970 to 1973, the aerobic digestion process has saved Metro Denver in excess of $500,000.

10.8 References

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2. Ashman, P. S., "Operating Experiences [illegible] Vacuum Filtration at St. Helens," *Water Pollut. Contr.* (1969), pp. 20-39.

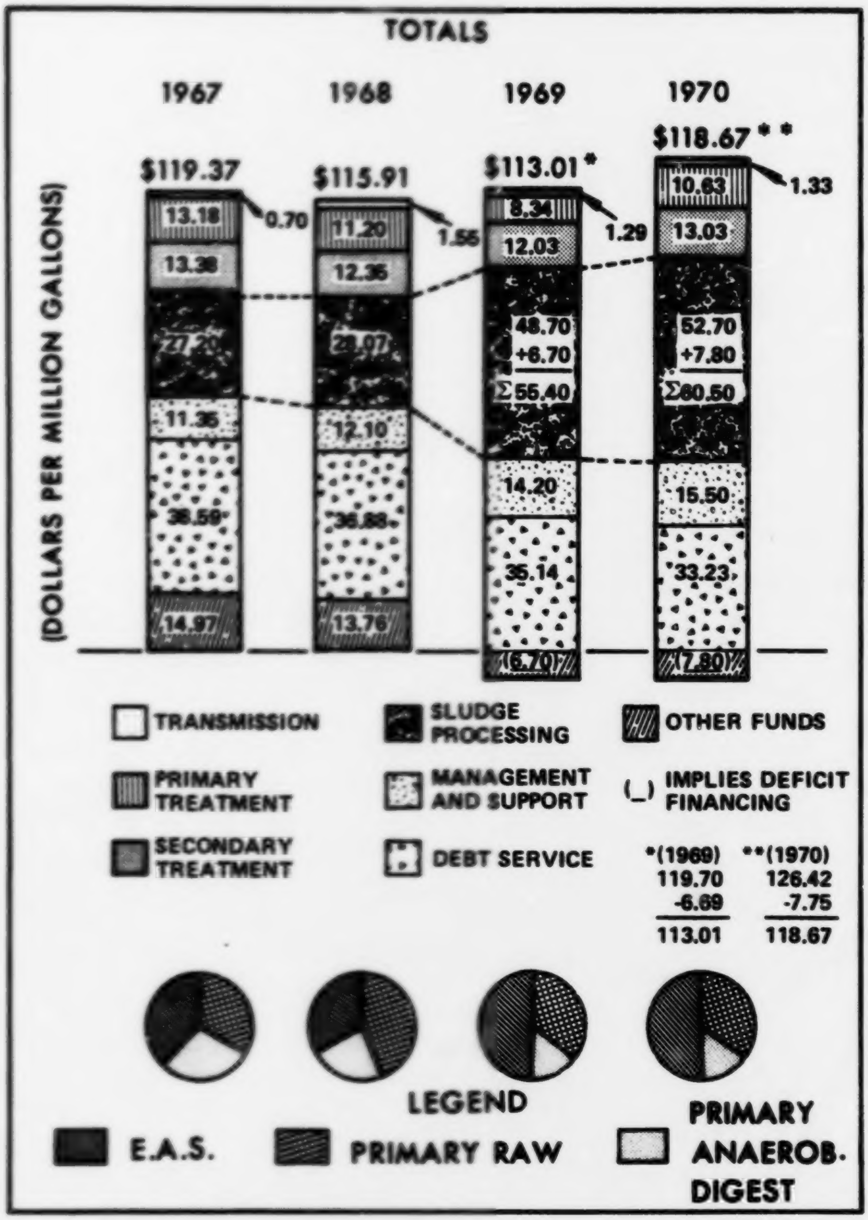

FIGURE 10-20. Unit costs of Metro wastewater treatment [9].

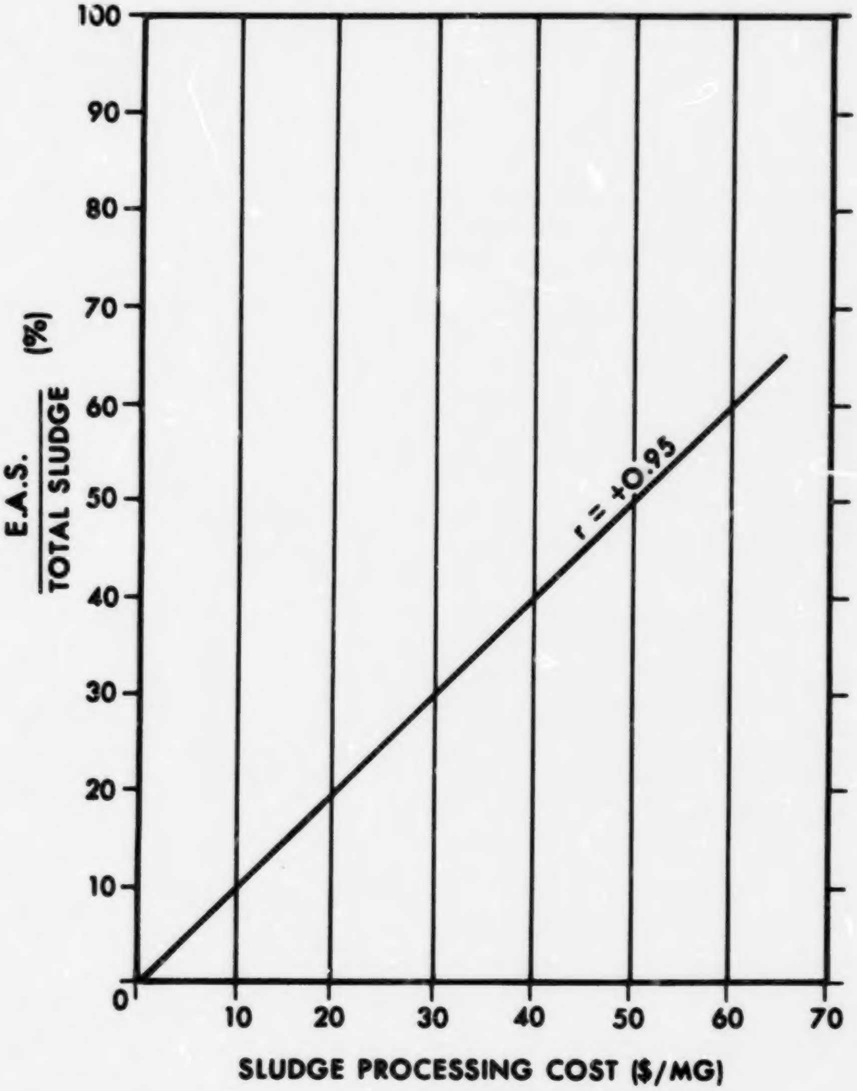

FIGURE 10-21. Sludge processing costs vs. EAS/total sludge produced (1968–1972) [9].

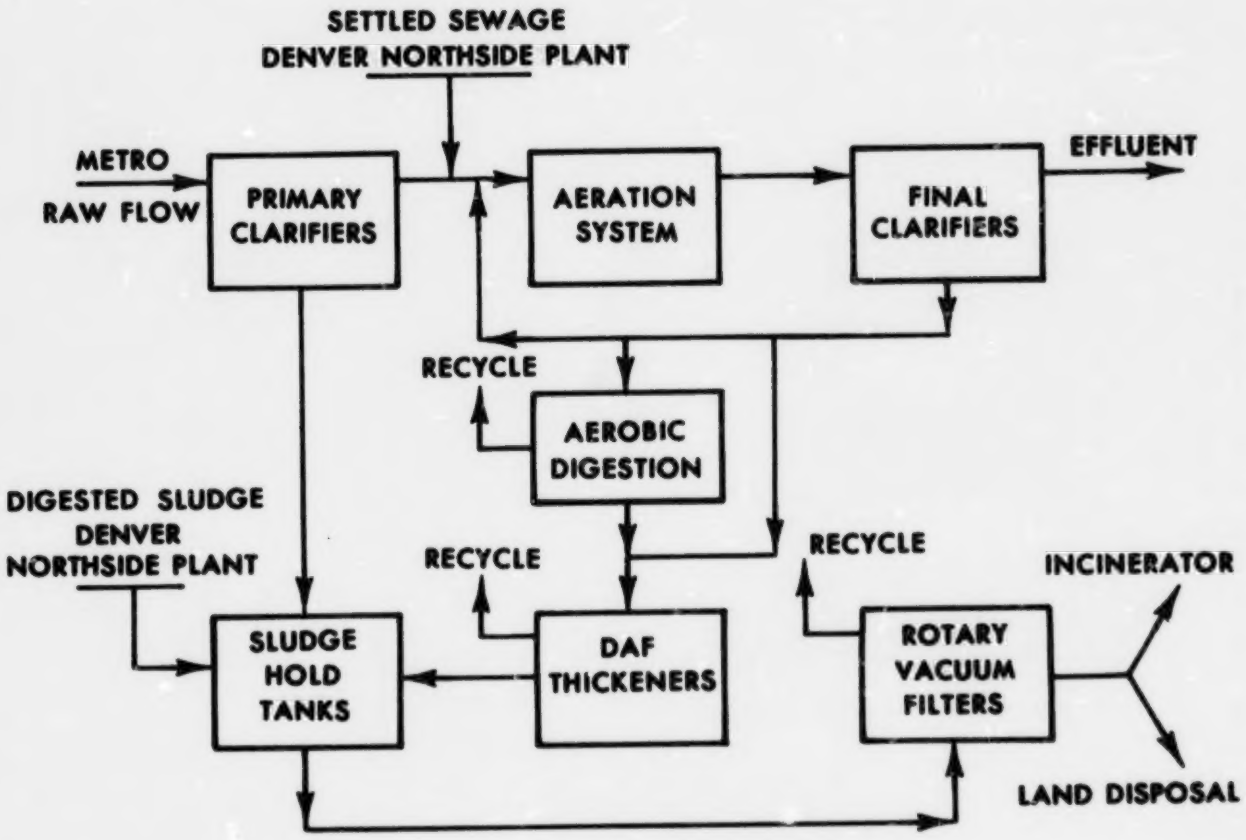
SETTLED SEWAGE
DENVER NORTHSIDE PLANT
METRO
RAW FLOW
PRIMARY CLARIFIERS
AERATION SYSTEM
FINAL CLARIFIERS
EFFLUENT
RECYCLE
AEROBIC DIGESTION
DIGESTED SLUDGE
DENVER
NORTHSIDE PLANT
RECYCLE
RECYCLE
INCINERATOR
SLUDGE HOLD TANKS
DAF THICKENERS
ROTARY VACUUM FILTERS
LAND DISPOSAL

FIGURE 10-22. Modified Metro Denver system.

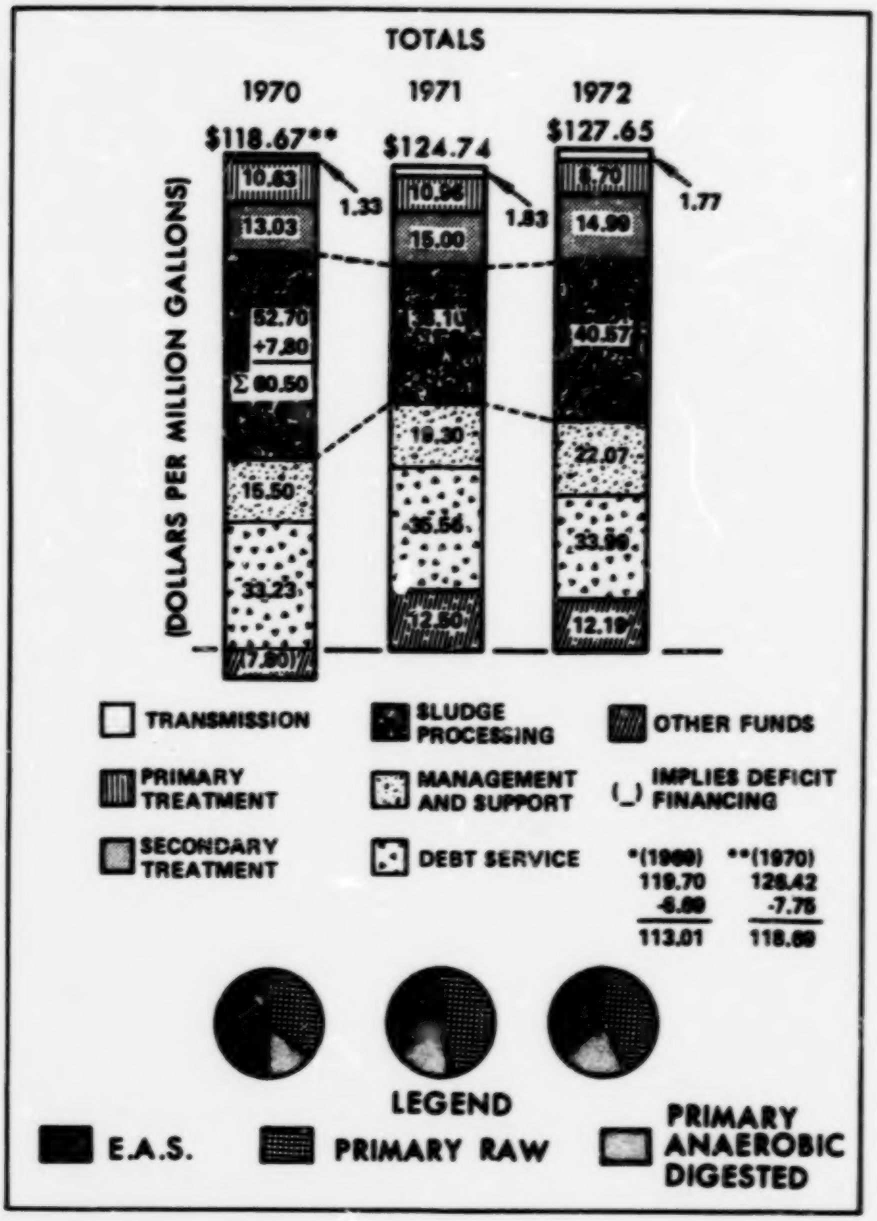

FIGURE 10-23. Unit costs of Metro wastewater treatment from 1970 to 1972 [9].

3. Ashman, P. S. and Roberts, P. F., "Operating Experiences with Vacuum Filtration at St. Helens: A Solution to the Problem," *Water Pollut. Contr.* (1970), pp. 638-648.

4. Stanbridge, H. H., "Operation and Performance of the Hogsmill Valley Sewage Treatment Works of the Greater London Council, 1958-1966." *Water Pollut. Contr.*, **67** (1), 21 (1968).

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7. Robson, C. M., Block, C. S., Nickerson, G. L., and Klinger, R. C., "Operational Experience of a Commercial Oxygen Activated Sludge Plant." Presented at WPCF, Atlanta, Georgia, Oct. 1972.

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CHAPTER 11

CASE HISTORIES OF SLUDGE TREATMENT BY HIGH TEMPERATURE AND PRESSURE

11.1 Heat Treatment

Heat treatment processes have been used for several years in Europe and the United States. The primary objective of heat treatment is to improve a sludge's dewaterability, while a secondary objective is to stabilize the sludge. The available proprietary systems include the Porteous, Zimpro LPO, and Farrer processes.

Porteous Process

The schematic diagram for a typical Porteous process is given in Figure 11-1. Heat is applied for elevating the sludge temperature from 350° to 400° F, and the pressure is raised to 150 to 300 psig. Steam is generally injected into the sludge, and this is followed by a sludge/water/sludge heat exchange system as shown in the diagram. Air injection is not normally practiced. Basic components of this system include sludge storage, grinding, a preheater, high pressure and temperature reactor, decanter thickener, auxiliary liquid treatment, off gas deodorizer, and a steam boiler. Colorado Springs, Colorado, is the only U.S. installation of this process that has been operating for any significant length of time. Two other locations commenced operation recently, and several more are planned. The process has evolved from a batch system to a continuous system.

Zimpro LPO Process

Figure 11-2 shows a schematic for a typical Zimpro LPO system. The principal differences between the Zimpro LPO process and the Porteous process are that air is added to improve heat exchange characteristics and fuel consumption, and a sludge to sludge heat exchanger is employed. About 26 United States installations of this system are now in existence, and most of these have been constructed within the past several years. Several more Zimpro LPO installations are under construction or planned.

Farrer Heat Treatment System

A schematic of the Farrer system is given in Figure 11-3. The Farrer system is similar to the continuous Porteous type process. At this time, there are few systems in operation in the United States.

11.2 Process Considerations

It is only within the past several years that significant United States operating and cost data on heat treatment processes have become available. Further, results from the United

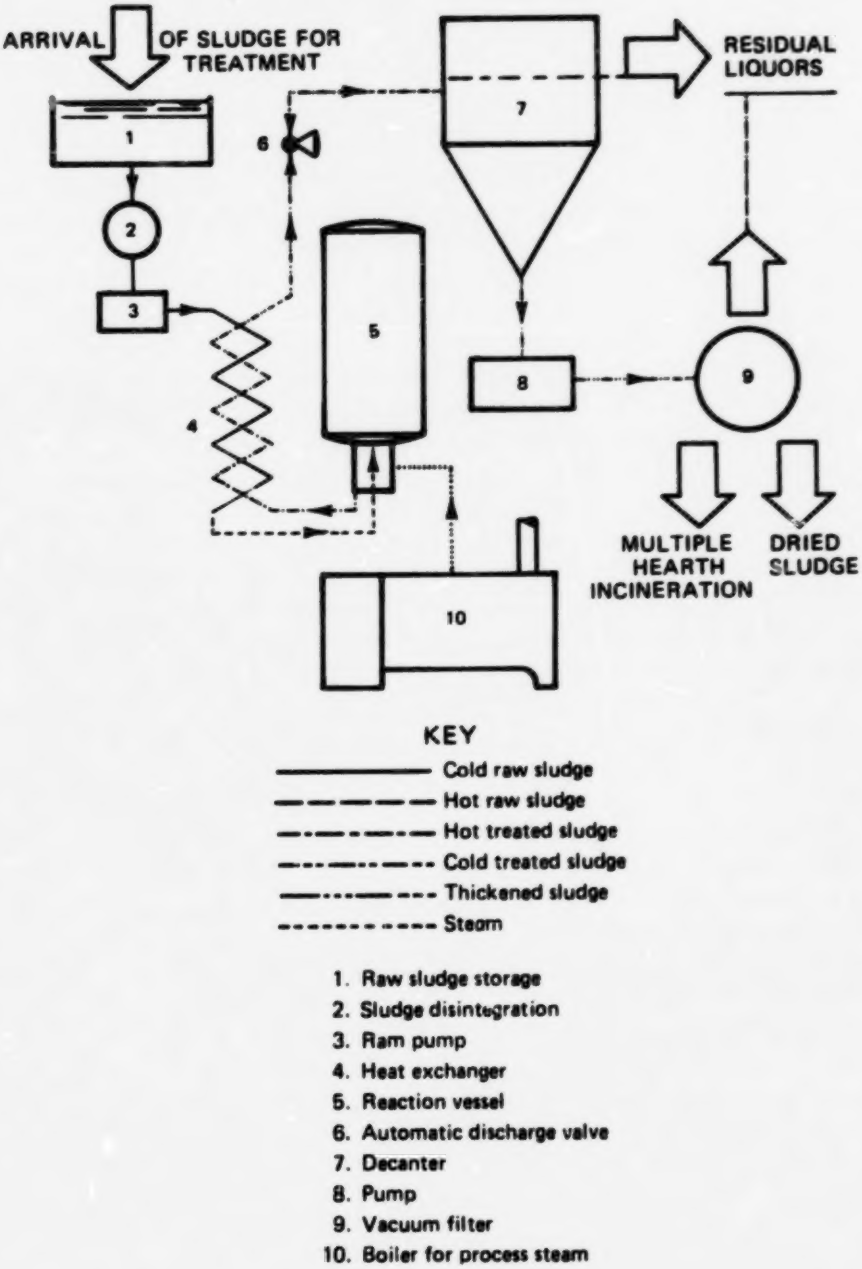
ARRIVAL OF SLUDGE FOR TREATMENT
RESIDUAL LIQUORS
1
2
3
4
5
6
7
8
9
10
MULTIPLE HEARTH INCINERATION
DRIED SLUDGE
KEY
Cold raw sludge
Hot raw sludge
Hot treated sludge
Cold treated sludge
Thickened sludge
Steam
1. Raw sludge storage
2. Sludge disintegration
3. Ram pump
4. Heat exchanger
5. Reaction vessel
6. Automatic discharge valve
7. Decanter
8. Pump
9. Vacuum filter
10. Boiler for process steam

FIGURE 11-1. Porteous process.

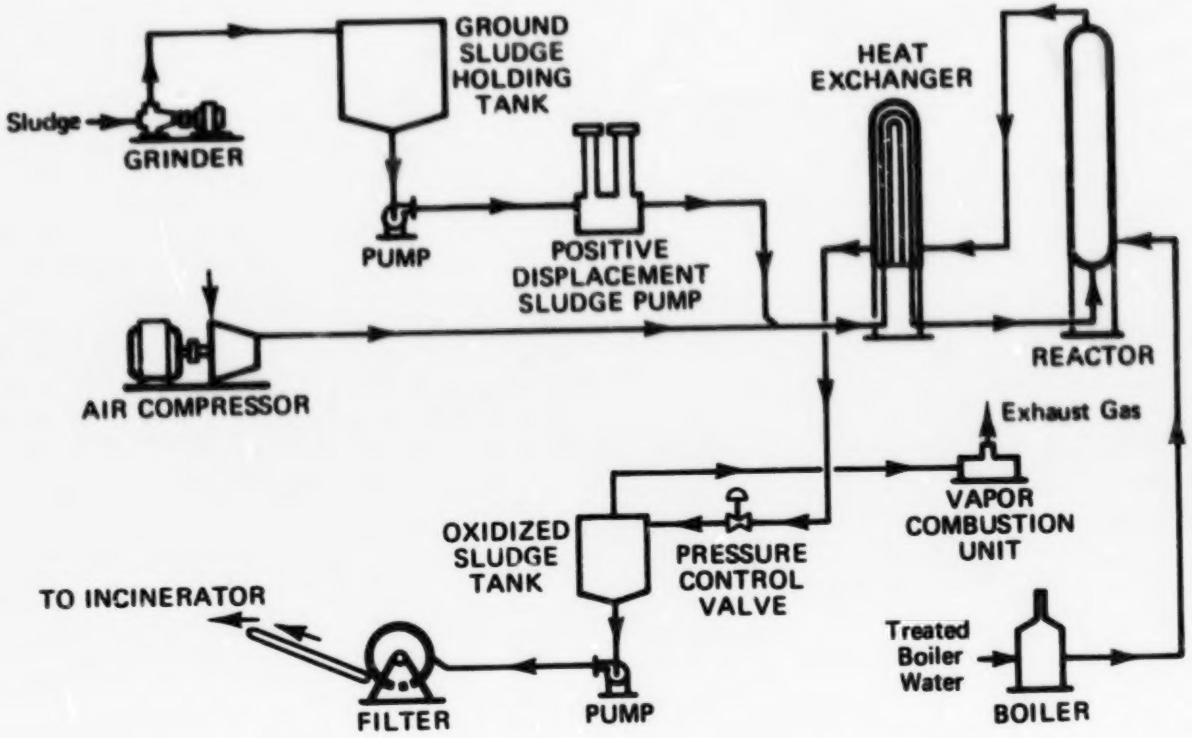
Sludge
GRINDER
GROUND SLUDGE HOLDING TANK
PUMP
POSITIVE DISPLACEMENT SLUDGE PUMP
HEAT EXCHANGER
REACTOR
AIR COMPRESSOR
Exhaust Gas
VAPOR COMBUSTION UNIT
OXIDIZED SLUDGE TANK
PRESSURE CONTROL VALVE
TO INCINERATOR
FILTER
PUMP
Treated Boiler Water
BOILER

FIGURE 11-2. Zimpro LPO system.

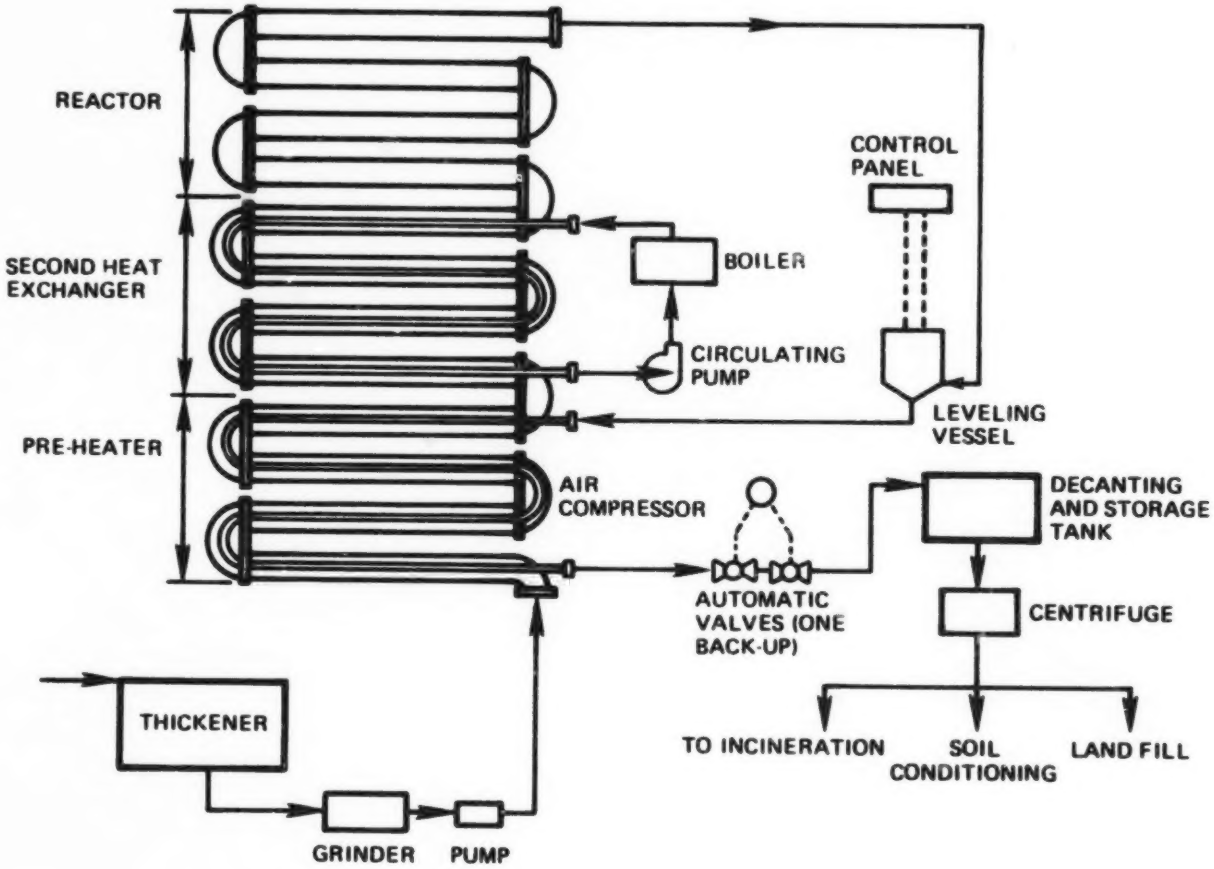
REACTOR
SECOND HEAT EXCHANGER
PRE-HEATER
CONTROL PANEL
BOILER
CIRCULATING PUMP
LEVELING VESSEL
AIR COMPRESSOR
AUTOMATIC VALVES (ONE BACK-UP)
DECANTING AND STORAGE TANK
CENTRIFUGE
THICKENER
GRINDER
PUMP
TO INCINERATION
SOIL CONDITIONING
LAND FILL

FIGURE 11-3. Farrer process system.

Kingdom are now in the technical journals. Difficulties with plants in the United Kingdom are generally attributed to the problems of maintaining such items of equipment as high pressure pumps, compressors, and high temperature and pressure reaction systems. Another difficulty has been in providing adequate treatment for the heat treated sludge cooking liquor. Plant difficulties in the United Kingdom were in some cases attributed to the installation of systems at older plants. A new plant with interrelated liquid and sludge heat treatment facilities might not have had troubles. However, some of the plants that have ceased operation were specifically designed with new liquid treatment facilities which could accommodate the heat treatment system recirculation loads. The principal cause of process cost and effluent quality problems appears to be a much higher degree of sludge solubilization with heat treatment than was predicted. Available information indicates high costs of operation, maintenance, and effluent quality problems are associated with heat treatment systems. Several U.S. plants have ceased operating heat treatment systems due to these problems, including Coors-Golden, Colorado; Santee, California; and Chattanooga, Tennessee. Several additional installations of heat treatment are in the planning, design, and construction phase.

11.3 Coors-Golden, Colorado

The Coors-Golden plant is an activated sludge system with a capacity of 5 mgd. The plant treated a combination of domestic and brewery wastes and installed a Porteous type heat treatment system in 1970. Coors Brewery utilizes its industrial engineering and treatment plant personnel to improve process selection and efficiencies. The heat treatment system was discontinued after about one year of operation due to the very high cooking liquor recirculation load as well as corrosion and other operating and cost problems. The cooking liquor from the sludge heat treatment system sometimes had a solids content as high as 20,000 mg/l. Even after heat conditioning the sludge fed to the belt type vacuum filters still required as much as 3.8 percent ferric chloride conditioning for good dewaterability. A burnt coffee odor persisted around the entire plant during heat treatment.

11.4 Colorado Springs, Colorado

Until the fall of 1973, Colorado Springs had a trickling filter plant, which removed 66 percent of the influent BOD. Plant flows were averaging 21 to 25 mgd with the actual design capacity being 18 mgd. The influent wastewater contained some industrial waste and had a BOD of about 300 mg/l and a suspended solids concentration of about 345 mg/l. A Porteous unit was installed during 1968 to 1969, and it was designed to treat 2,000 lb/hr of sludge at an operating temperature of 370° F and a pressure of 250 psi. Figure 11-4 gives the flow diagram of the process.

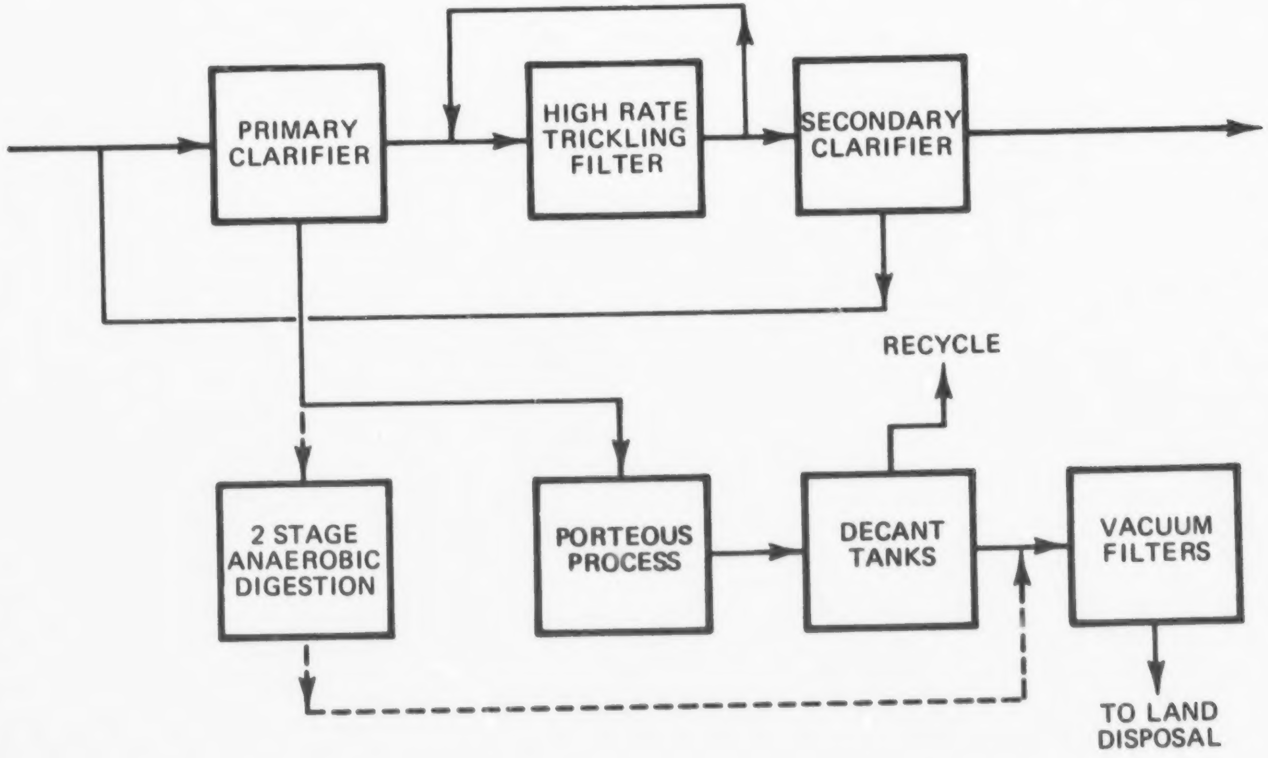
PRIMARY CLARIFIER
HIGH RATE TRICKLING FILTER
SECONDARY CLARIFIER
RECYCLE
2 STAGE ANAEROBIC DIGESTION
PORTEOUS PROCESS
DECANT TANKS
VACUUM FILTERS
TO LAND DISPOSAL

FIGURE 11-4. Flow diagram of Colorado Springs with heat treatment.

Sherwood and Phillips [1] reported vacuum filtration rates of 12 $lb/hr/ft^2$ and concentrations of 37 percent after heat treatment. No chemical conditioning was required. While prior to installation of heat treatment, chemical costs were $18 to $20 per ton for ferric chloride and lime. Filtrate and decant streams were handled with no additional aeration equipment required. No material balance data were given. Periodic visits to the plant at the times for which the cost data were reported revealed that the operating personnel were having problems with the recycle load from heat treatment and were receiving odor complaints. The recycle load was greater than expected even though this was a primary and trickling filter process plant sludge rather than an activated sludge. Plant process work was carried out to reduce the recycle load and this included lime precipitation of the liquor. Kochera [2] published data on an additional year of operation. He stated that the chemical conditioning costs prior to heat treatment had averaged from $20 to $40 per ton, while the operating costs for the Porteous system were averaging $15 a ton (fuel, power, labor, and water). It was not clear whether the maintenance costs were included in these figures, and apparently no allowance was made for any costs in treating the recirculation load. Recent data [3] indicate that the average Porteous system, vacuum filtration, and land disposal costs for 1972 were $30/ton of dry sludge solids. The Porteous system alone was approximately $22 per ton and did not include recirculation liquor treatment costs. Vacuum filtration costs, therefore, were about $5/ton, and about $2.50 to $3.00/ton were required for land disposal. New activated sludge and Porteous systems were started up in late 1973 at Colorado Springs. Little data are available on their operation. Table 11-1 summarizes the operating costs for the Porteous system and vacuum filter at Colorado Springs.

TABLE 11-1

COLORADO SPRINGS SLUDGE PROCESSING COST

Cost Cited	$/ton	Reference Source
Operating–Porteous/V.F.	2	[1]
Operating–Porteous/V.F.	15	[2]
Operating/Maintenance–Porteous/V.F./Land	30	[3]

11.5 Borough of Pudsey–United Kingdom

The Pudsey plant installed a Farrer process in the 1969 to 1970 period for treatment of a sludge with about 82 percent trickling filter humus and 18 percent activated sludge [4]. The sludge handling system at Pudsey is shown in Figure 11-5. The heat treatment

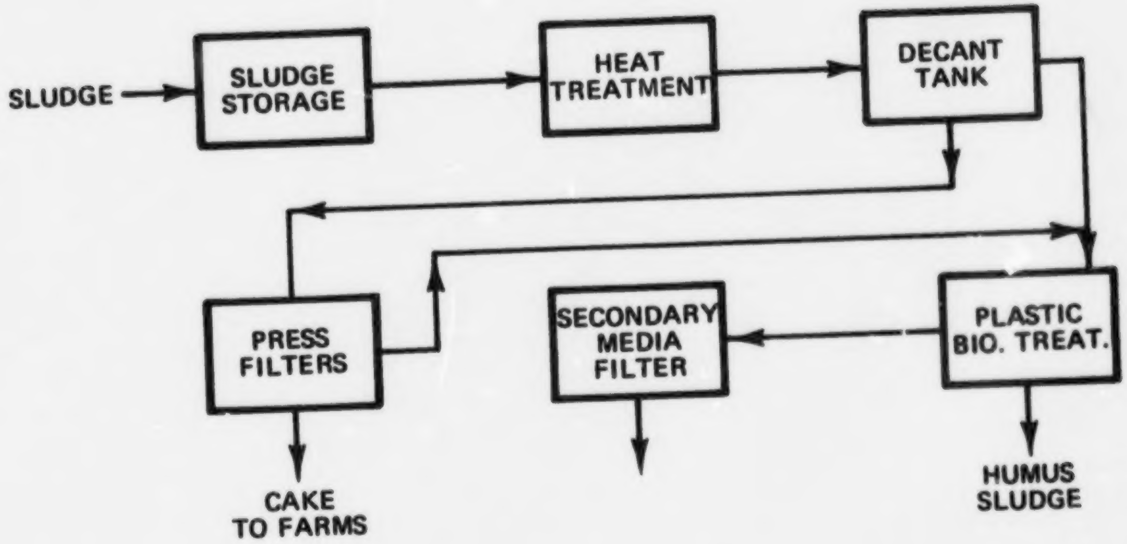
SLUDGE
SLUDGE STORAGE
HEAT TREATMENT
DECANT TANK
PRESS FILTERS
SECONDARY MEDIA FILTER
PLASTIC BIO. TREAT.
CAKE TO FARMS
HUMUS SLUDGE

FIGURE 11-5. Pudsey sludge system [4].

process included provision of a decant tank, press filters to dewater the heat treated sludge, and disposal of the dewatered sludge cake to land. A separate treatment system was used for the cooking liquor and involved a plastic media trickling filter and a secondary media filter. Successful operation of this system at the Pudsey plant was not possible in 1½ years, and severe operating and maintenance problems were encountered. Sufficient experience was gained to make it possible to estimate costs. The estimated cost for heat treatment and dewatering was $37.20/ton of dry solids, of which $22.32 was for operation and maintenance and $14.88 was for capital. The cost of treating the recycled liquors from heat treatment by plastic trickling filter and obtaining a 50 percent BOD reduction was estimated to be about $5/ton. Thus the total cost of heat treatment exclusive of filter cake disposal was $42.20/ton of dry solids.

In general, Great Britain has found heat treatment extremely effective in improving filtration characteristics. However, it recognizes that approximately half of the solids go into solution and need to be treated. If treatment is accomplished by a biological method, more sludge is created [5].

11.6 Kalamazoo, Michigan

A Zimpro LPO unit is in operation at Kalamazoo, Michigan. The wastes entering the plant are a combination of domestic, paper mill, and pharmaceutical wastes. The influent to the Kalamazoo plant, however, is essentially made up of 18 mgd of municipal waste and 12 mgd of paper mill wastes. The Kalamazoo plant was, until 1966, a primary treatment plant with anaerobic digestion and drying bed treatment of sludge. In 1967, some secondary capacity was provided and the sludge was lagoon treated. When the lagoons became a problem, mechanical sludge dewatering was introduced. The 34 mgd Kalamazoo treatment plant currently has a high rate activated sludge process. This high rate activated sludge system was designed for shock loading and treatment of up to 100,000 lb/day of BOD.

The initial testing program at Kalamazoo which led to the selection of the Zimpro LPO unit was performed with both lagoon sludge and sludge to be expected in the future. The Zimpro system was installed for sludge processing in combination with multiple hearth incineration [6]. Figure 11-6 shows a schematic of the Kalamazoo sludge handling system. The heat treatment system consists of three Zimpro units designed to process a total of 97.5 tons/day of sludge. This system could readily handle the sludge from approximately a 45 mgd facility. The units were designed to run at a temperature of 360° F and a pressure of 315 to 325 psig. Retention time in the heat treatment system is about one hour, and demineralized water at a pH of 9.0 is required. A heat recovery system was included in the design but has been inoperable due to the dirty off gases from the incinerator. With an operable method for heat recovery, up to 800,000 cu ft of gas per month could be saved. However, as noted in the section on incineration, when particularly dry filter cake is fed to a multiple hearth incinerator, poor quality off gases

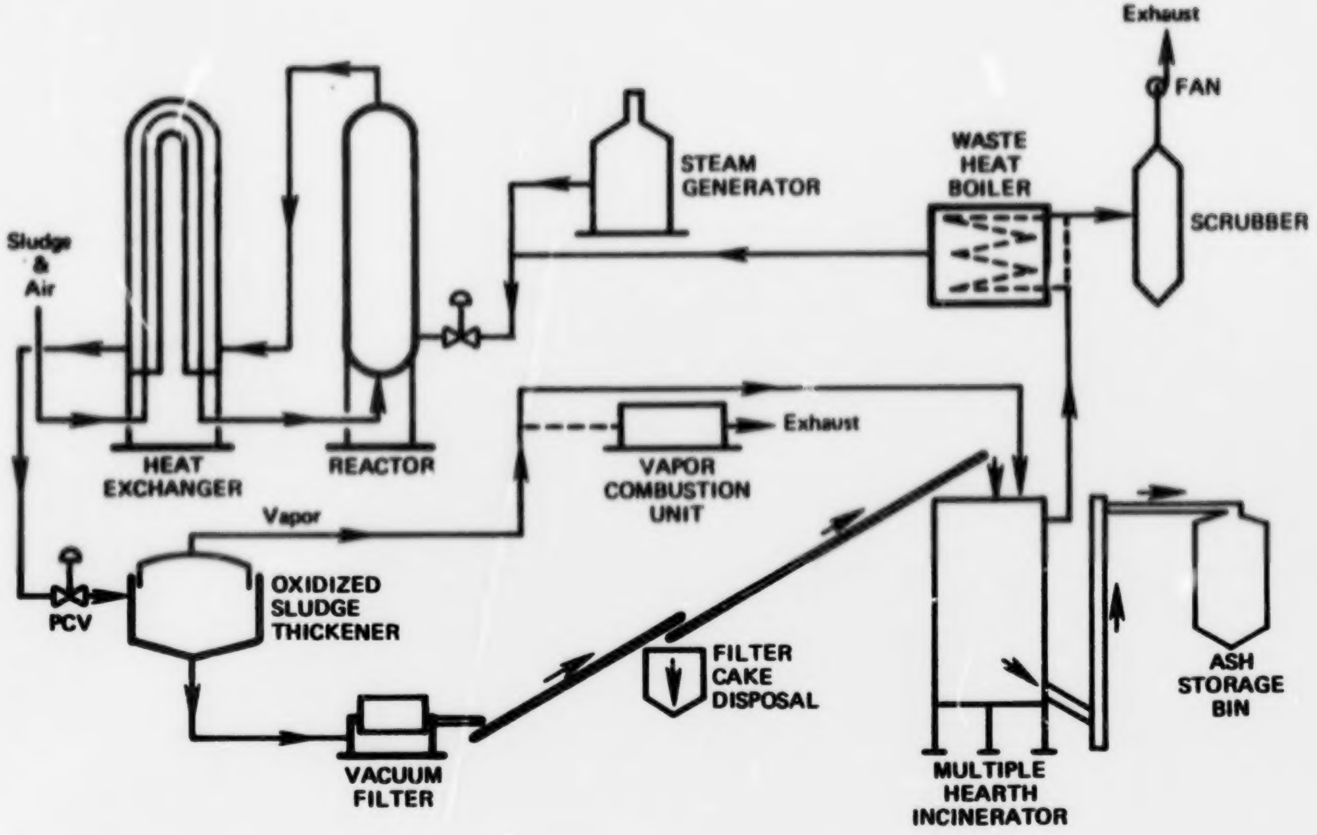
Exhaust
FAN
Sludge & Air
STEAM GENERATOR
WASTE HEAT BOILER
SCRUBBER
HEAT EXCHANGER
REACTOR
Exhaust
VAPOR COMBUSTION UNIT
Vapor
PCV
OXIDIZED SLUDGE THICKENER
FILTER CAKE DISPOSAL
VACUUM FILTER
MULTIPLE HEARTH INCINERATOR
ASH STORAGE BIN

FIGURE 11-6. Kalamazoo, Michigan, sludge disposal facilities [7].

result. Operation of the Kalamazoo plant's sludge thickening and dewatering processes has been quite good. The sludge thickener provided a sludge with a solids concentration of 9.7 percent. The vacuum filter had a production rate of 4.9 lb/hr/sq ft and yielded a cake with a solids concentration of 45 percent [6]. Good incinerator operation resulted. No data were available on the decantate from the oxidized sludge thickener. The most recent cost data (1972) on sludge processing at Kalamazoo are presented in Table 11-2.

TABLE 11-2

KALAMAZOO SLUDGE PROCESSING COSTS

Sludge Process	Cost $/ton
Thickening	2.20
Heat Treatment	7.52
Dewatering	5.43
Incineration	6.48
Amortization	10.00
	31.63

This information gives a cost for thickening, heat treatment, dewatering, incineration, and amortization of $31.63/ton. This cost does not include an estimate of the capital, operating, and maintenance costs associated with treatment of the recycle streams from heat treatment. In December of 1973, Kalamazoo plant personnel reported that the recirculation load from heat treatment increases the BOD loading on the secondary treatment system by 35 to 40 percent. Despite the fact that the aeration system at Kalamazoo is designed for a high shock loading, this recirculation load has resulted in some problems. Low plant BOD and suspended solids removals have occurred. BOD removal efficiency at Kalamazoo has averaged 75 to 80 percent and the effluent contains 60 to 130 mg/l of BOD. Suspended solids removal has averaged 70 to 80 percent and the effluent contains 50 to 88 mg/l suspended solids.

11.7 Ft. Lauderdale, Florida—Plant A

Plant A in Ft. Lauderdale is a new 8.5 mgd activated sludge plant. A schematic of the plant's sludge handling facility is shown in Figure 11-7. As can be seen, primary and excess activated sludge is gravity thickened and then pumped to heat treatment. Heat treated sludge goes first to a hold tank and then to a solid bowl scroll-type centrifuge for dewatering. The aeration system at Ft. Lauderdale was overdesigned by 44 percent to accommodate a predicted 9,000 lb/day BOD load in the primary effluent as well as a 4,000 lb/day BOD load anticipated from cooking liquor recirculation. This plant started up in December, 1971, and a series of heat exchangers failed primarily because of corrosion problems. An oversize exchanger was temporarily installed, and this provided approximately one year of operation. This operation was with less than adequate facilities and under conditions which resulted in excessive fuel consumption. During this period of operation the sludge thickening and dewatering processes worked poorly. The heat treated sludge thickened to only 2 to 3 percent, and the solids capture during dewatering was less than 50 percent. A buildup of fine solids occurred throughout the system. A new titanium heat exchanger was started up in January of 1973. As of July 19, 1973, the same poor sludge thickening and dewatering results were being experienced. Fine solids carry-over into the final clarifiers and odor problems were still encountered. The principal problem has been maintaining an adequate sludge solids removal rate through the heat treatment and dewatering systems. Remedial measures tried have included use of the fill and draw method of thickener operation, use of flocculants in the thickening step, and the addition of chlorine in the thickener to improve overflow clarity. Table 11-3 illustrates solids capture being experienced at Ft. Lauderdale during July, 1973.

TABLE 11-3

TOTAL SOLIDS—SLUDGE AND CENTRATE

Date 1973	Thickened to Heat Treatment	To Centrifuge	Centrate	Cake
	%			
July 5	3.1	2.6	1.2	42.2
July 7	3.0	2.4	1.0	40.6
July 11	2.4	2.4	1.2	33.2
July 16	1.8	1.8	1.3	28.8

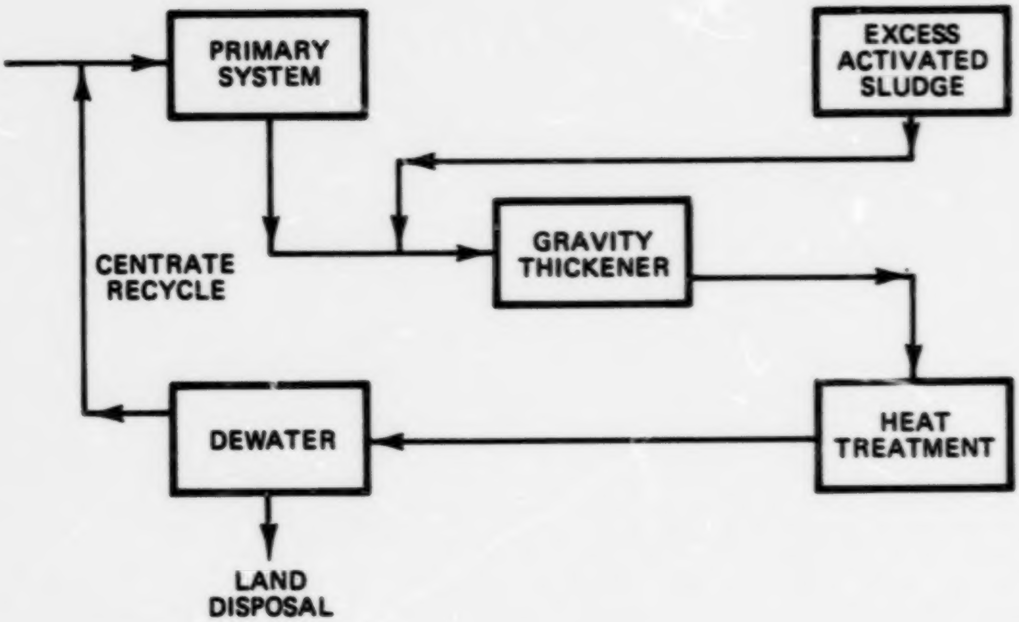
PRIMARY SYSTEM
EXCESS ACTIVATED SLUDGE
GRAVITY THICKENER
CENTRATE RECYCLE
DEWATER
HEAT TREATMENT
LAND DISPOSAL

FIGURE 11-7. Ft. Lauderdale sludge handling system.

Less than 50 percent total solids capture in the overall heat treatment dewatering steps still prevailed.

Old Plant A was located on the same site and employed conventional treatment methods. Treatment costs amounted to $186.47 per million gallons in the last year of its operation (1971). Costs are not reported for 1972 because of abnormalities due to equipment, operating, and maintenance problems in New Plant A. The cost in 1973, pertaining to the new plant, is somewhat high ($224.84 per million gallons). It is still difficult at the time to evaluate effects of abnormal and other inflationary conditions on costs.

11.8 References

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3. Personal communication with plant manager, Colorado Springs, Colorado, 1973.
4. Hirst, G., Mulhall, K. G., and Hemming, M. L., "The Sludge Heat Treatment Plant at Pudsey." Northeastern Branch of the Institute of Water Pollution Control (Mar. 25, 1971).
5. Pickford, J. (ed.), "Sludge Treatment and Disposal." Fourth Public Health Engineering Conference Proceedings, Department of Civil Engineering, Loughborough University, Jan. 1971.
6. Swets, D. H., Pratt, L, and Metcalf, C., "Combined Industrial-Municipal Thermal Sludge Conditioning and Multiple Hearth Incineration." WPCF Annual Meeting, Atlanta, Georgia, 1972.
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